PB84-101898

Review of New Source Performance Standards for Petroleum Refinery Claus Sulfur Recovery Plants

(U.S.) Environmental Protection Agency
Research Triangle Park, NC

Aug 83

U.S. Department of Commerce
National Technical Information Service
NTIS

PB84-101898

United States
Environmental Protection
Agency

Office of Air Quality
Planning and Standards
Research Triangle Park NC 27711

EPA-450/3-83-014
August 1983

Air

EPA

Review of New Source Performance Standards for Petroleum Refinery Claus Sulfur Recovery Plants

NSPS

REPRODUCED BY
NATIONAL TECHNICAL
INFORMATION SERVICE
U.S. DEPARTMENT OF COMMERCE
SPRINGFIELD, VA. 22161

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO.
EPA 450/3-83-014

2.

3. RECIPIENT'S ACCESSION NO.
PB84 101898

4. TITLE AND SUBTITLE
Review of New Source Performance Standards for Petroluem Refinery Claus Sulfur Recovery Plants

5. REPORT DATE
August 1983

6. PERFORMING ORGANIZATION CODE

7. AUTHOR(S)

8. PERFORMING ORGANIZATION REPORT NO.

9. PERFORMING ORGANIZATION NAME AND ADDRESS
Office of Air Quality Planning and Standards
U. S. Environmental Protection Agency
Research Triangle Park, North Carolina 27711

10. PROGRAM ELEMENT NO.

11. CONTRACT/GRANT NO.

12. SPONSORING AGENCY NAME AND ADDRESS
DAA for Air Quality Planning and Standards
Office of Air, Noise, and Radiation
U. S. Environmental Protection Agency
Research Triangle Park, North Carolina 27711

13. TYPE OF REPORT AND PERIOD COVERED

14. SPONSORING AGENCY CODE
EPA/200/04

15. SUPPLEMENTARY NOTES

16. ABSTRACT

This document provides background information on sulfur emissions and emissions control for claus sulfur recovery plants in petroleum refineries. State and Federal emission regulations are summarized. The claus process is described with emphasis on factors which affect emissions. Emission controls are also detailed with accompanying costs and performance data. Other environmental and energy impacts associated with claus emissions controls are outlined. Detailed energy balances and operating cost estimates are presented in appendices for two representative control systems. This information was developed for a four-year review of the New Source Performance Standard for refinery claus sulfur plants (40CFR60, Sub part J) as required by the Clean Air Act Amendments of 1977.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution Pollution Control Standards of Performance Claus Sulfur Plants Tail Gas Treaters Petroleum Refineries	Air Pollution Control Sulfur Recovery Plants	13 B

18. DISTRIBUTION STATEMENT Release unlimited	19. SECURITY CLASS *(This Report)* unclassified	21. NO. OF PAGES 122
	20. SECURITY CLASS *(This page)* unclassified	22. PRICE

EPA Form 2220-1 (9-73)

EPA-450/3-83-014

Review of New Source Performance Standards for Petroleum Refinery Claus Sulfur Recovery Plants

Emission Standards and Engineering Division

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Air, Noise, and Radiation
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

August 1983

This report has been reviewed by the Emission Standards and Engineering Division of the Office of Air Quality Planning and Standards, EPA, and approved for publication. Mention of trade names or commercial products is not intended to constitute endorsement or recommendation for use. Copies of this report are available through the Library Services Office (MD-35), U. S. Environmental Protection Agency, Research Triangle Park, N.C. 27711, or from National Technical Information Services, 5285 Port Royal Road, Springfield, Virginia 22161.

TABLE OF CONTENTS

Page

1. SUMMARY 1-1

1.1 CONTROL TECHNOLOGY 1-1
1.2 ECONOMIC CONSIDERATIONS AFFECTING THE NSPS 1-1
1.3 OTHER FINDINGS 1-1

2. INTRODUCTION 2-1

2.1 NSPS AND NSPS REVIEW 2-1
2.2 BACKGROUND INFORMATION 2-1
2.3 SULFUR RECOVERY IN REFINERIES 2-2
2.4 REFINERY SULFUR PLANT STATISTICS 2-7
2.5 SELECTION OF SULFUR PLANTS FOR NSPS CONTROL 2-8
2.6 REFERENCES 2-8

3. CURRENT STANDARDS FOR REFINERY SULFUR PLANTS 3-1

3.1 AFFECTED FACILITIES 3-1
3.2 CONTROLLED POLLUTANTS AND EMISSION LEVELS 3-1
3.3 STATE REGULATIONS 3-1
3.4 TESTING AND MONITORING REQUIREMENTS 3-3
3.5 REFERENCES 3-4

4. STATUS OF CONTROL TECHNOLOGY 4-1

4.1 EXTENDED CLAUS REACTION PROCESSES 4-1
4.2 TAIL GAS SCRUBBING PROCESSES 4-5
4.3 COMMERCIAL STATUS OF EMISSION CONTROLS 4-13
4.4 REFERENCES 4-19

5. COMPLIANCE STATUS OF REFINERY SULFUR PLANTS 5-1

5.1 AFFECTED FACILITIES 5-1
5.2 COMPLIANCE TEST RESULTS 5-1
5.3 OPERABILITY OF NSPS UNITS 5-3
5.4 STATUS OF EMISSION MONITORS 5-5
5.5 EMISSION TESTING 5-6
5.6 REFERENCES 5-6

6. MODEL PLANTS AND COST ANALYSES 6-1

6.1 MODEL PLANTS 6-1
6.2 CONTROL LEVELS 6-5
6.3 COST ANALYSIS 5-5
6.4 REFERENCES 6-18

TABLE OF CONTENTS (continued)

Page

7. OTHER IMPACTS REVIEWED 7-1

7.1 NON-AIR ENVIRONMENTAL IMPACTS 7-1
7.2 ENERGY AND ENERGY-RELATED IMPACTS 7-2
7.3 OTHER IMAPCTS 7-3
7.4 REFERENCES 7-4

8. RECOMMENDATIONS 8-1

8.1 REVISIONS TO NSPS 8-1
8.2 REVISIONS TO MONITORING REQUIREMENTS 8-3
8.3 REVISIONS TO COMPLIANCE TESTING REQUIREMENTS 8-3
8.4 SUMMARY OF RECOMMENDATIONS 8-4
8.5 REFERENCES 8-4

1. SUMMARY

1.1 CONTROL TECHNOLOGY

Based on emissions data obtained in the original NSPS study and recently obtained emissions and reliability data from an industry survey, the most effective emission control technology for refinery Claus sulfur plants are systems capable of achieving 99.9 percent overall sulfur recovery. These systems cost as much as the parent Claus plant, but have shown good reliability and have been successfully integrated into refinery operations at 70 sites, with another 19 planned or under construction These systems include SO_2 scrubbing (Wellman-Lord), reduction-Stretford sulfur recovery (Beavon), and reduction-amine absorption (SCOT, ARCO, and BSRP/MDEA). All systems subject to the NSPS levels of 250 ppmv SO_2 or 300 ppmv total sulfur have successfully complied to date.

1.2 ECONOMIC CONSIDERATIONS AFFECTING THE NSPS

The primary issue involving review of the NSPS is the cost of controls. To determine cost trends, facilities of 10.16, 50.8, and 101.6 megagrams per day (Mg/D) were modelled. At 10.16 Mg/D, the cost-effectiveness of control was assessed at 2,125 dollars per megagram of sulfur dioxide (SO_2) removed. At 50.8 and 101.6 Mg/D, the corresponding cost-effectiveness indeces were found to be $880/Mg and $675/Mg, respectively. The current NSPS would then require a maximum expenditure of about $1,430/Mg (at the 20.32 Mg/D cutoff), but more typically would be considerably less than $900/Mg SO_2 based on current and planned sulfur plant capacities.

1.3 OTHER FINDINGS

No significant adverse environmental impacts were noted for the control technologies. Control systems energy consumption is significant and accounts for 5 to 13 percent of total sulfur plant operating costs, for the models examined.

For systems with tail gas incineration, EPA Method 6 and continuous SO_2 analyzers are used for initial compliance testing and monitoring, respectively. For systems without tail gas incineration, a modified EPA Method 15 has been used and possible changes to this method for measuring reduced sulfur compounds may be forthcoming. Continuous monitors for total reduced sulfur have recently been introduced and are currently being evaluated by the EPA. No satisfactory hydrogen sulfide (H_2S) monitors have been identified.

2. INTRODUCTION

2.1 NSPS AND NSPS REVIEW

The United States Environmental Protection Agency (EPA) proposed new source performance standards for petroleum refinery sulfur plants under Section 111 of the Clean Air Act on October 4, 1976, (41FR43866). These regulations were promulgated on March 15, 1978, (43FR10866) and amended on October 25, 1979, (44FR61542). The regulations applied to Claus sulfur recovery plants greater than 20 long tons per day (LT/D) capacity, the construction or modification of which commenced after October 4, 1976.

The Clean Air Act Amendments of 1977 require that the Administrator of the EPA review and, if appropriate, revise established standards of performance for new stationary sources at least every 4 years. The purpose of this report is to review and assess the need for revision of the existing standards for refinery sulfur plants based on developments that have occurred or are expected to occur within the petroleum refining industry. The information presented in this report was obtained from reference literature, discussions with industry representatives, trade organizations, control equipment vendors, EPA regional offices, and State and local agencies.

2.2 BACKGROUND INFORMATION[1]

Petroleum refineries convert naturally occuring "crude" petroleum liquids into marketable fuels such as heating oil and gasoline in a number of chemical processes. During this processing, impurities such as sulfur are liberated as gaseous hydrogen sulfide (H_2S) and are collected with plant gases known as process or fuel gas. To satisfy air pollution regulations which effectively limit the sulfur in fuel gas, and to reduce corrosion problems, refineries "sweeten" or remove hydrogen sulfide from the fuel gas before burning it in process heaters and boilers.

Sweetening processes currently used in petroleum refineries consist of scrubbing the sour gases with liquids which preferentially absorb hydrogen sulfide and carbon dioxide over other species. Regeneration of the scrubbing solutions evolves a secondary gas stream containing concentrated hydrogen sulfide with lesser amounts of carbon dioxide, water vapor, and hydrocarbons.

Refinery process water may also contain dissolved gases such as ammonia and H_2S, which require removal before the water may be reused or discharged. The water is subjected to thermal or steam stripping which liberates the dissolved gases into a gas stream consisting of water vapor, hydrogen sulfide, hydrocarbons, and ammonia.

In many instances, the choice of disposition of this gas stream is to route it to sulfur recovery with other H_2S-rich streams. Alternatively, the sour water stripper overhead may be incinerated where sulfur dioxide regulations permit.

2.3 SULFUR RECOVERY IN REFINERIES

At one time, many refineries sold the H_2S-rich gas streams to neighboring chemical plants, or "scavengers", as feedstock for sulfuric acid or elemental sulfur production. Recent trends, however, are to convert the H_2S on-site to marketable liquid sulfur via the Claus process.

2.3.1 Claus Process[2]

Figure 2-1 is a representative process diagram of the Claus process. Basically, the overall chemical reaction is a thermal and catalytic oxidation of H_2S to elemental sulfur in the gaseous phase:

(1) $H_2S + 1/2\ O_2 \rightarrow H_2O + S$

The reaction is exothermic in that considerable heat is generated by the Claus process. Additionally, one mole of water vapor and one mole of sulfur vapor are formed for each mole H_2S converted.

Actual Claus reactions occur in stages as shown in Figure 2-1.

The sour gases are initially combusted in a furnace where sufficient air is admitted to convert one-third of the H_2S to SO_2:

(2) $H_2S + 3/2\ O_2 \rightarrow SO_2 + H_2O$

Then the remaining 2/3 H_2S and the 1/3 SO_2 react:

(3) $2H_2S + SO_2 \rightarrow 3S + 2H_2O$

Combining reactions (2) and (3) yields the overall Claus reaction (1).

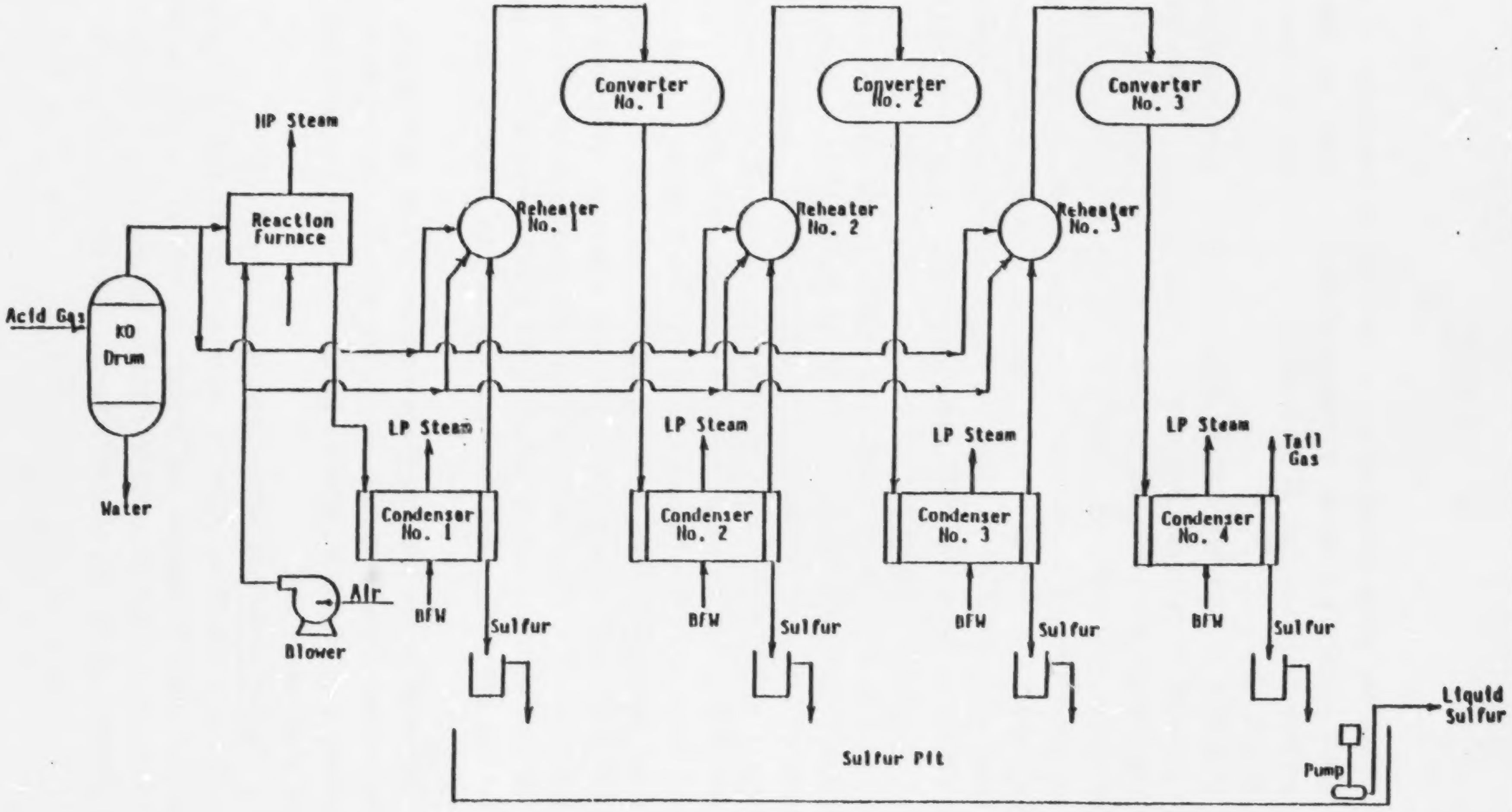

Acid Gas
KO Drum
Water
HP Steam
Reaction Furnace
Air
Blower
Reheater No. 1
Converter No. 1
Reheater No. 2
Converter No. 2
Reheater No. 3
Converter No. 3
LP Steam
Condenser No. 1
BFW
Sulfur
LP Steam
Condenser No. 2
BFW
Sulfur
LP Steam
Condenser No. 3
BFW
Sulfur
LP Steam
Tail Gas
Condenser No. 4
BFW
Sulfur
Sulfur Pit
Pump
Liquid Sulfur

Figure 2-1. Flow diagram for a three-stage Claus sulfur recovery facility

Since the above reactions are exothermic, the conversion of H_2S to elemental sulfur is promoted by removal of heat via shell and tube heat exchangers; therefore, the Claus plant is a net exporter of steam as well as sulfur.

Reaction (1), in addition to being favored by lower temperatures, is also promoted by catalysts and removal of sulfur vapor. Therefore, upon leaving the furnace (where up to 60 percent of the Claus reaction has taken place), the gases are subjected to successive catalytic stages and sulfur condensers, with each successive catalytic stage operated at lower temperatures. In lieu of emission regulations, the Claus plant is normally operated with two or three catalytic stages, depending on economic considerations, with the final condenser outlet routed to an incinerator.

2.3.2 Claus Plant Emissions

The only significant source of emissions is the Claus incinerator; fugitive sulfur emissions are possible due to leaks and atmospheric venting of liquid sulfur storage and transfer areas. Emissions are typically sulfur dioxide where incinerators are operated at temperatures of 650°-800°C, sufficient to destruct sulfides and elemental sulfur vapor. Lower oxidizer temperatures of 540-650°C may be adequate to destruct gaseous sulfides where the sulfide concentration has been significantly reduced upstream by tail gas treating. Emissions are a direct function of the Claus conversion efficiency, which will be discussed in the next section. For a typical Claus plant operating at 96 percent conversion efficiency, emissions are 8 percent by weight of the incoming sulfur feed.

Other emissions from the Claus incinerator are small amounts of hydrocarbons, nitrogen oxides, and carbon monoxide, all of which are dependent upon fuel combustion parameters and generally unrelated to Claus plant operation.

2.3.3 Factors Affecting Sulfur Dioxide Emissions[3,4,5]

Design of the Claus plant is important, as the type of catalyst, number of catalytic stages, and process controls all influence emissions. Obviously, the number of catalytic stages determines to a great extent the ultimate sulfur recovery efficiency. A Claus furnace may operate at

60 percent conversion, while successive catalytic stages may increase conversion to 85-90 percent for one, 92-95 percent for two, and 96-97 percent for three stages. The type of catalyst is also important, as newer alumina catalysts show 1 to 2 percent improvement over the conventional bauxite catalysts. Finally, the Claus plant requires both upstream monitoring of acid gas feed and downstream monitoring of tail gas sulfur species to enable operation at optimum conditions.

Claus plant operation is heavily influenced by the feedstock composition. The presence of hydrocarbons, carbon dioxide, and ammonia all adversely affect Claus plant performance, first by the dilution of reactive H_2S and SO_2 in the Claus plant, but more importantly by adverse side reactions. Hydrocarbons and ammonia if not properly combusted, form solid compounds which rapidly degrade catalyst surfaces and Claus performance. Carbon dioxide also reacts with hydrogen sulfide, thereby diminishing sulfur recovery:

(4) $CO_2 + H_2S \rightarrow H_2O + COS$

(5) $COS + H_2S \rightarrow H_2O + CS_2$

Thus, two additional sulfur compounds, carbonyl sulfide (COS) and carbon disulfide (CS_2) are formed in the Claus furnace and, though hydrolyzed in the subsequent catalytic stages, are significant contributors to Claus emissions.

Hydrocarbons may also react in the Claus furnace to form CS_2:

(6) $CH_4 + 2S_2 \rightarrow CS_2 + 2H_2S$

Operator control of the process is the most influential factor affecting emissions. In order to maximize sulfur conversion, the following parameters must be controlled:

- stoichiometric ratio of H_2S to SO_2
- furnace, catalyst bed, and condenser temperatures
- catalyst activity

Figure 2-2 illustrates the importance of maintaining the H_2S-SO_2 ratio at 2 to 1. This is accomplished by metering the air flow to the furnace to convert exactly one-third of incoming H_2S to SO_2. Air control is complicated by variable feedstock flow rates and changes in composition, both of which affect furnace stoichiometry. If air to the furnace is deficient, the H_2S-SO_2 ratio is too high and sulfur recovery diminishes;

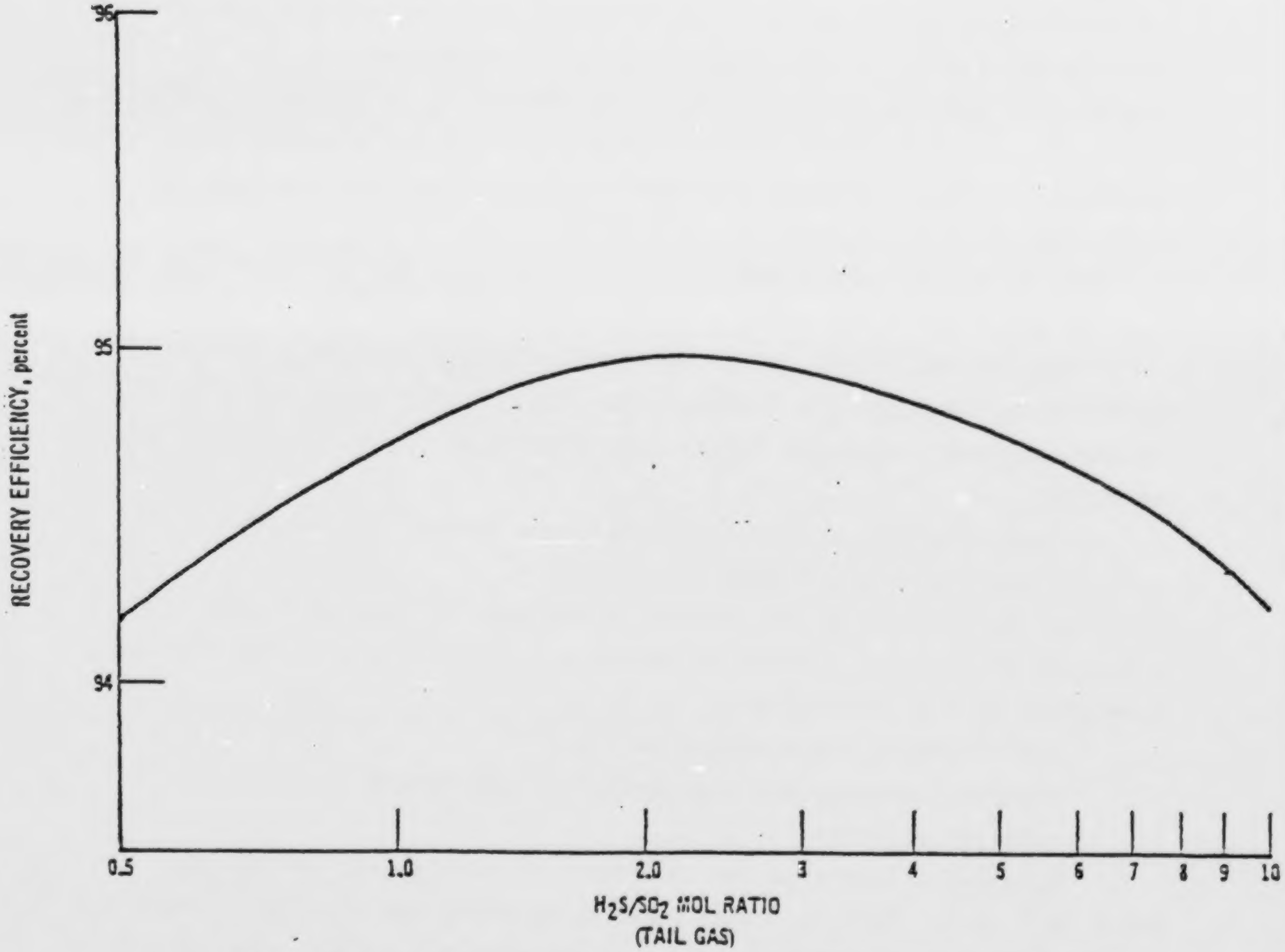


Figure 2-2. Theoretical Claus sulfur recovery efficiency vs. Mole Ratio.

if air is excessive, too much SO_2 is formed, the ratio becomes less than 2 to 1, and recovery again diminishes.

Temperatures must be maintained at optimum levels; high temperatures decrease reaction equilibrium and sulfur condensation while low temperatures may promote adverse reactions on catalyst surfaces. The final condenser must especially be maintained at a low temperature to minimize sulfur vapor losses.

Catalyst activity is maintained by periodic regeneration or replacement, which require either a period of suboptimum operation or plant shutdown.

Operation of a Claus plant at low loads may adversely affect performance. One vendor reported a 2 to 3 percent loss in recovery at 20 percent load. Operation from two-thirds capacity up to 120 percent capacity is reported with no loss in recovery.

2.4 REFINERY SULFUR PLANT STATISTICS[6,7,8,9,10,11]

In 1973, total Claus sulfur capacity in U.S. refineries totalled 8,000 megagrams per day (Mg/D). 1974 construction was estimated at over 1,000 Mg/D. Since statistics have not been kept on whether the growth since 1973 has been due to new facilities or replacements, the actual Claus capacity is not known, but is considerably greater than 10,000 Mg/D. Recent construction announcements show that for 1981, nine sulfur plants were installed totalling 800 Mg/D, with a tenth plant of unspecified capacity constructed. In 1982, eight plants having 516 Mg/D were scheduled for completion, with two others of unspecified size due to start up.

Vendor announcements indicate that at least 13 new Claus facilities will be constructed in 1983, totalling 2,009 Mg/D capacity (See Table 4-2). Construction announcements in Hydrocarbon Processing for early 1983 project that 28 new Claus plants will be constructed in the 1983-85 time frame, 25 of which will total 5,184 Mg/D. Of these, 19 individual plants totalling 5,083 Mg/D will be sized greater than 20.32 Mg/D capacity. Six plants of 101 Mg/D total capacity will be constructed that are not subject to Federal NSPS.

These figures indicate that strong growth in sulfur plant construction will continue, the average size unit will be large (~200 Mg/D), and the total capacity of units not covered by NSPS will account for less than 2 percent of new plant growth.

2.5 SELECTION OF SULFUR PLANTS FOR NSPS CONTROL

Refinery sulfur plants were originally selected for NSPS development because of their potential for emissions of sulfur dioxide in significant quantities. Though the actual emissions from Claus plants have likely decreased significantly from the estimated 306,715 megagrams annually in 1973[11] due to replacements with NSPS units and considerable retrofitting of existing units, the potential for emissions from Claus plants without controls remains. For example, a 101.6 Mg/D plant operating at 96 percent conversion for 350 days per year at rated capacity could emit 2,845 megagrams per year sulfur dioxide, a criteria pollutant.

The widespread use of emission controls on Claus plants, hereafter referred to as "tail gas units", on many retrofitted existing Claus plants and practically all refinery Claus plants installed since 1975, indicates that the technology for Claus emissions control is well established and generally accepted by industry. Therefore, the ingredients for NSPS development--growth, emission potential, and demonstrated control technology--that were present prior to development of the NSPS, persist at this time.

2.6 REFERENCES

1. Standards Support and Environmental Impact Statement Volume 1: Proposed Standards of Performance for Petroleum Refinery Sulfur Recovery Plants, EPA 450/2-76-016a, September 1976, pp. 3.1-3.2.

2. Reference 1, pp. 3.2-3.9.

3. Reference 2.

4. Parnell, David C., "Differences in Design of Claus Plants for Various Applications", Paper Number 22d, Spring National AIChE Meeting, April 9, 1981.

5. GPA Panelist Outlines Claus Process Improvements in Sulfur Recovery, Oil & Gas Journal, p. 9299, August 7, 1978.

6. Reference 1, pp. 3.1-3.2.

7. "HPI Construction Boxscore", Hydrocarbon Processing, October 1981, pp. 3-18.

8. Letter, W. T. Knowles, Shell Oil Company to Charles B. Sedman, U.S. EPA, August 24, 1982.

9. Letter, M. A. Peterson, Union Oil Co. of California, to C. Sedman, U.S. EPA, September 15, 1982.

10. Letter, J. C. Brocoff, Ralph M. Parsons Co., to S. T. Cuffe, U.S. EPA, dated February 16, 1983.

11. "HPI Construction Boxscore," Hydrocarbon Processing, February 1983.

12. Reference 1.

3. CURRENT STANDARDS FOR REFINERY SULFUR PLANTS

3.1 AFFECTED FACILITIES

Existing new source performance standards (NSPS) for new, modified, and reconstructed refinery sulfur recovery facilities limit sulfur emissions from Claus sulfur recovery plants of greater than 20.32 megagrams per day (Mg/D) capacity. A Claus sulfur recovery plant is defined as a "process unit which recovers sulfur from hydrogen sulfide by a vapor-phase catalytic reaction of sulfur dioxide and hydrogen sulfide".[1]

3.2 CONTROLLED POLLUTANTS AND EMISSION LEVELS

The NSPS limits emissions of reduced sulfur compounds, hydrogen sulfide, and sulfur dioxide as follows:

Reduced Sulfur Compounds

Reduced sulfur compounds from Claus plants are defined as hydrogen sulfide, carbonyl sulfide, and carbon disulfide. These are limited to 0.030 percent (300 ppmv) by volume at zero percent oxygen on a dry basis. These are measured only if the emission control system is a reduction system not followed by an incinerator. This is roughly equivalent to 99.8-99.9 percent sulfur recovery.

Hydrogen Sulfide

Hydrogen sulfide emissions are limited to 0.0010 percent (10 ppmv) by volume at zero percent oxygen on a dry basis. Hydrogen sulfide measurements are required only if the emission control system is a reduction system not followed by an incinerator.

Sulfur Dioxide

Sulfur dioxide emissions are limited to 0.025 percent (250 ppmv) by volume at zero percent oxygen on a dry basis if emissions are controlled by an oxidation control system or a reduction control system followed by incineration. This is comparable to the 99.8-99.9 percent control level for reduced sulfur.

3.3 STATE REGULATIONS

In 1976, when NSPS were proposed, most States having petroleum refineries generally required 99 percent sulfur removal for new Claus plants.[2] The Environment Reporter reveals some recent changes, but in general, the States having the majority of refineries still require 99 percent sulfur recovery (equivalent to about 1300 ppmv SO_2 at stack conditions).[3] Table 3-1 summarizes selected 1972 and 1982 standards for refinery sulfur plants. One noticeable omission is for California which has standards set by local air pollution control districts. (One district having refineries generally requires control equivalent to the NSPS.) Hydrogen sulfide regulations were generally based on ground level concentrations. The listing in Table 3-1 may understate the ultimate control requirements, as other State regulations such as best available control technology (BACT) or prevention of significant deterioration (PSD) mandates may well supercede emission codes.[4]

Table 3-1. SELECTED STATE REGULATIONS FOR NEW SULFUR RECOVERY PLANTS AT 101.6 Mg/D

State	1972	1982
Delaware	2000 ppmv (98.5%)	Process Wt. (93.4%)
Illinois	--	2000 ppmv (98.5%)
Louisiana	.01 lb/lb S input (99%)	.01 lb/lb S input (99%)
New Jersey	15000 ppmv (90%)	15000 ppmv (90%)
Ohio	.01 lb/lb S input (99%)	Process Wt. (99.2-99.4% for 101.6 Mg/D)
Oklahoma	.01 lb/lb S input (99%)	.01 lb/lb S input (99%)
Pennsylvania	Process Wt. (98.4%)	500 ppmv (99.6%)
Texas*	Process Wt. (87.6%)	Process Wt. (2200 ppmv or 98.4%)

*In most instances superceded by BACT requirements (Reference 4).

Since most refineries are located in industrialized urban areas, and because essentially all sulfur plants potentially emit greater than 90.74 megagrams per year and are subject to additional regulations such as BACT/PSD mentioned above, essentially all sulfur plants installed within the last 5 years have been required to install tail gas treaters. The only exceptions have been small sulfur plants in rural areas. States contacted generally require tail gas treaters as best available control technology (BACT) unless the source is shown to have a negligible impact on air quality.[5,6]

3.4 TESTING AND MONITORING REQUIREMENTS

3.4.1 Testing Requirements

Performance tests to verify compliance with the standards for refinery sulfur plants must be conducted within 60 days after achieving full capacity operation, but not later than 180 days after the initial startup of the facility. This is a uniform requirement for all affected facilities under 40 CFR 60.8. The EPA reference methods to be used in connection with the affected facilities include:

1. Method 4 for moisture content
2. Method 6 for SO_2
3. Method 15 for H_2S and reduced sulfur compounds

For Method 6, a series of three runs each spanning a minimum of four consecutive hours is required. For Method 15, three runs each consisting of 16 samples taken over a minimum of three hours is required. Reference Method 4 is conducted simultaneously with Method 15, sampling at a rate proportional to the gas velocity for a minimum of four continuous hours sampling for each run.

Total reduced sulfur is expressed as SO_2 equivalent under Method 15 by the following formula:

$$SO_2 \text{ equivalent} = \Sigma\ (H_2S, COS, 2CS_2)d$$

where: SO_2 equivalent = the sum of the concentration of each of the measured compounds expressed as sulfur dioxide in ppm

H_2S = hydrogen sulfide, ppm

COS = carbonyl sulfide, ppm

CS_2 = carbon disulfide, ppm

d = dilution factor, dimensionless

$$\text{Average } SO_2 \text{ equivalent} = \frac{\sum_{i=1}^{N} SO_2 \text{ equivalent } i}{N(1-B_{wo})} \qquad 3\text{-}1$$

where: average SO_2 equivalent = average SO_2 equivalent in ppm, dry basis as

SO_2 equivalent = SO_2 in ppm as determined in equation 3-1

N = Number of analyses performed

B_{wo} = Fraction of volume of water vapor in the gas stream as determined by Method 4

3.4.2 Monitoring Requirements

A continuous monitoring system is required under the NSPS to monitor and record the concentration of SO_2 or alternatively, reduced sulfur and H_2S compounds, on Claus tail gas exhaust to the atmosphere. Specifications for continuous sulfur dioxide monitors were promulgated in Appendix B, 40 CFR Part 60.

3.5 REFERENCES

1. Federal Register, Wednesday, March 15, 1978, Part III 10866-10873.

2. Standard Support and Environmental Impact Statement Volume 1: Proposed Standards of Performance for Petroleum Refinery Sulfur Recovery Plants, EPA 450/2-76-016a, September 1976, pp. 3.13-3.15.

3. Environment Reporter, State Air Laws, Bureau of National Affairs (updated to 7/9/82), pp. 201:001-556:0523.

4. Letter from Sam Crowther, Texas Air Control Board, to S.T. Cuffe, U.S. EPA, dated January 7, 1983.

5. Telephone Conversation: C. Sedman, EPA, to Sam Crowther, Texas Air Control Board, March 16, 1982.

6. Telephone Conversation: C. Sedman, EPA, to Jim Stone, Louisiana Bureau of Environmental Services, March 17, 1982.

4. STATUS OF CONTROL TECHNOLOGY

The total sulfur emissions from a Claus sulfur plant were established in Chapter 2 as a direct function of the extent to which the Claus reaction reaches completion. Thermodynamically, the Claus reaction is limited at normal operating temperatures and pressures to 97-98 percent recovery, but in actual practice is reduced by process limitations such as unsteady state operation and catalyst aging.[1] Therefore, to reduce emissions to the atmosphere, the Claus process must be augmented by (1) extending the Claus reaction into a lower temperature liquid phase, or (2) adding a scrubbing process to the Claus exhaust stream.

4.1 EXTENDED CLAUS REACTION PROCESSES

There are at least five processes currently available to augment or extend the Claus reaction beyond the recoveries normally achieved in a conventional Claus with three catalytic stages. These are the BSR/Selectox, Sulfreen, Cold Bed Absorption, Maxisulf, and IFP-1 processes. Of these four, the only domestic refinery applications to date involve the IFP-1 process; therefore, only the IFP-1 will be discussed in detail. The other processes are briefly described herein as applicable.

4.1.1 BSR Selectox[2,3]

The BSR/Selectox I process, recently developed by Union Oil of California and the Ralph M. Parsons Company, is designed to provide a sulfur recovery efficiency in the range of 99 percent, in conjunction with a three-stage Claus.

The BSR/Selectox I is a fixed bed catalytic process consisting of two steps. In the first step, tail gas from the second stage of the Claus plant is heated to above 288°C (500°F) in a reducing gas generator fueled by substoichiometric air and refinery fuel gas. The hot gases are passed over a catalyst bed where all sulfur species are converted to hydrogen sulfide. The gas is cooled, reheated, and passed over a proprietary catalyst to oxidize the H_2S to elemental sulfur. Sulfur is condensed out with the remaining tail gas passed to the final Claus stage.

Close control of $H_2S:SO_2$ ratio in the Claus plant is not as critical as with Claus and other extended Claus reaction schemes. Up to 99 percent sulfur recovery is reported on an overseas refinery application.

4.1.2 Sulfreen[4,5]

The Sulfreen process converts H_2S and SO_2 contained in Claus tail gas to elemental sulfur at temperatures of 127°C to 150°C (260°F to 300°F) by extension of the Claus reaction.

Claus tail gas is first scrubbed with liquid to wash out entrained sulfur liquid and sulfur vapor. The tail gas is then introduced to a battery of reactors where the lower temperatures push the Claus reaction toward completion on the surfaces of a special alumina catalyst. A regeneration gas, usually nitrogen, periodically desorbs the sulfur-laden catalyst beds, first driving off water vapor and carbon dioxide at 300°C (572°F) and then sulfur at 400°C (752°F). The sulfur is condensed out of the carrier gas, the carrier gas scrubbed in a sulfur wash, and then returned to the regeneration cycle.

A Sulfreen unit may consist of as little as two reactors, one in absorption and one in desorption service. The gases from the reactors being desorbed are incinerated before discharge to the atmosphere.

4.1.3 Amoco CBA[6]

The cold bed adsorption (CBA) process, developed by Amoco Production Company, is essentially the same concept as the Sulfreen process, except low temperature acid gas feed is used as the regeneration gas. A recent study assesses the CBA capability on a two-stage Claus plant at 98 percent recovery. Currently, three units (one on a natural gas plant in the United States) are in operation with capacities from 15 to 900 metric tons of sulfur per day.

4.1.4 Maxisulf[7]

The Maxisulf process,recently developed by Davy McKee, is similar in principle to the Sulfreen and Amoco CBA processes and features a cyclic, two-reactor process, one absorbing and one desorbing. The key feature is that a heated slipstream of Claus tail gas is used for the desorbing gas, then recombined with tail gas, entering the absorbing reactor. Thus, a closed loop, forced circulation desorption scheme is avoided. Efficiencies of 99 percent on refinery application are cited by the vendor. Two units are scheduled for construction in Germany.

4.1.5 IFP-1[8,9]

The IFP-1 (Institut Francais du Petrole) process is the only Claus extension type of tail gas process to be successfully applied on U.S. refinery Claus plants. It was initially applied at two refineries in 1973 as a retrofit second-stage to one-stage Claus plants. Larger installations followed as shown later in Table 4-1.

The IFP-1 process is essentially a liquid-phase Claus reactor which accepts Claus tail gas directly with no conditioning. The reactor is a packed column with a specially designed "boot" for collecting liquid sulfur. Metal salts catalyze the reaction which takes place in a high boiling point solvent, polyethylenglycol (PEG), above the melting point of sulfur--in the range of 121-126°C (250-260°F). The metal salts form a complex with H_2S and SO_2 in the feed gas, which in turn reacts with additional H_2S and SO_2 to form elemental sulfur and regenerate the catalyst. Sulfur coalesces and settles into the boot of the reactor, from which it is drawn as a molten product.

Gas typically leaving the reactor contains about 1500-2500 ppmv sulfur which includes essentially all COS and CS_2 formed in the Claus plant, about 300 ppmv sulfur vapor (the equilibrium concentration of sulfur vapor at 126°C), and the unreacted H_2S and SO_2. Conversion efficiencies on a nonrefining application of 99.3 percent have been reported. The reactor exhaust containing 1500-2500 ppmv sulfur is incinerated before discharge to the atmosphere. This represents overall control of roughly 99.0 percent.

Conversion efficiencies are maximized by (1) operating the IFP at H_2S to SO_2 ratios of as near 2:1 as possible and (2) operating the first Claus reactor at a higher temperature than normal to minimize COS/CS_2 formation.

Operation slightly above the 2:1 H_2S to SO_2 ratio is practiced due to the adverse effects of operation below 2:1. When the Claus tail gas is deficient in H_2S to carry the Claus reaction toward completion, the IFP solvent evolves absorbed SO_2 which decreases efficiency and increases sulfur emissions. Operation at long periods under H_2S deficient conditions may result in deterioration of the solvent/catalyst complex, where emissions increase until the unit is shut down and IFP solvent regenerated or completely replaced.

Figure 4-1 illustrates the IFP-1 process.

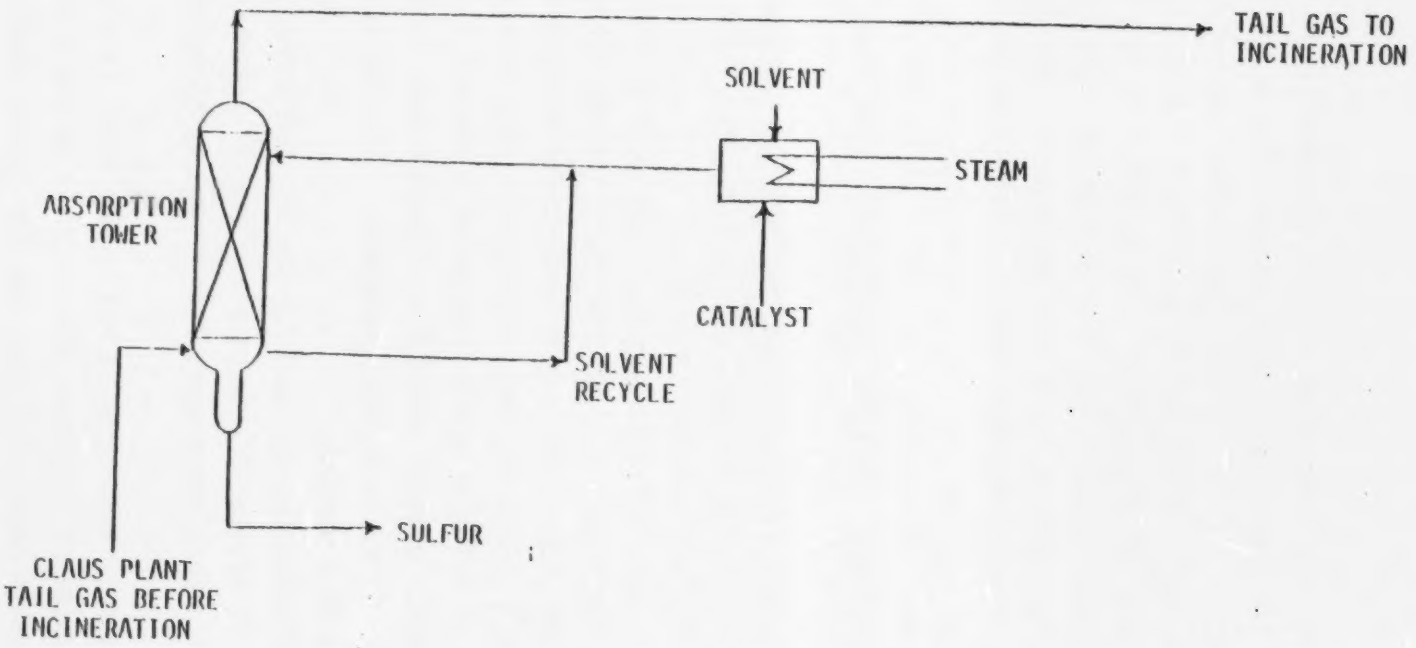

TAIL GAS TO INCINERATION
SOLVENT
STEAM
ABSORPTION TOWER
CATALYST
SOLVENT RECYCLE
SULFUR
CLAUS PLANT TAIL GAS BEFORE INCINERATION

Figure 4-1. Flow diagram for IFP-1 Claus tail-gas clean-up process.

4.2 TAIL GAS SCRUBBING PROCESSES

There are essentially two generic types of tail gas scrubbing processes--the first where Claus tail gas is oxidized and the oxidized sulfur (SO_2) absorbed by caustic scrubbing and the second where Claus tail gas is reduced, and the reduced sulfur (H_2S) absorbed by scrubbing with solvents or caustic reagents. Initially, the first tail gas scrubbers were mainly the sulfur dioxide/caustic type. Subsequently, the vast majority have been the reduction scrubber variety. For subsequent modelling and analyses, the reduction scrubber systems have been chosen as representative technologies. Both processes are described herein as demonstrated technologies.

4.2.1 Oxidation Tail Gas Scrubbers

At least three processes were developed to scrub SO_2 from incinerated Claus tail gas and recycle the concentrated SO_2 stream back to the Claus for conversion to elemental sulfur or, alternatively, send the concentrated SO_2 to a sulfuric acid plant. These were the Wellman-Lord, Stauffer Aquaclaus, and IFP-2. Since only the Wellman-Lord has been applied successfully to U.S. refineries, it is the only process of its type examined.

4.2.1.1 The Wellman-Lord Process[10,11] The Wellman-Lord process was developed by Wellman-Power Gas Incorporated and has been applied to various industrial SO_2 sources.

Figure 4-2 illustrates the Wellman-Lord process as applied to Claus tail gases. The Wellman-Lord system uses a wet regenerative process to reduce stack gas sulfur dioxide concentration to less than 250 ppmv or approximately 99.9 percent sulfur recovery.

Claus plant tail gas is incinerated and all sulfur species are oxidized to sulfur dioxide. Gases are then cooled and water quenched to remove excess water and lower gas temperatures to absorber conditions. The SO_2-rich gas is then contacted countercurrently with a solution of sodium sulfite (Na_2SO_3) and sodium bisulfite ($NaHSO_3$) which reacts with the SO_2 to form the bisulfite:

$$SO_2 + Na_2SO_3 + H_2O \rightarrow 2NaHSO_3$$

The off-gas is reheated (where required) and vented to the atmosphere.

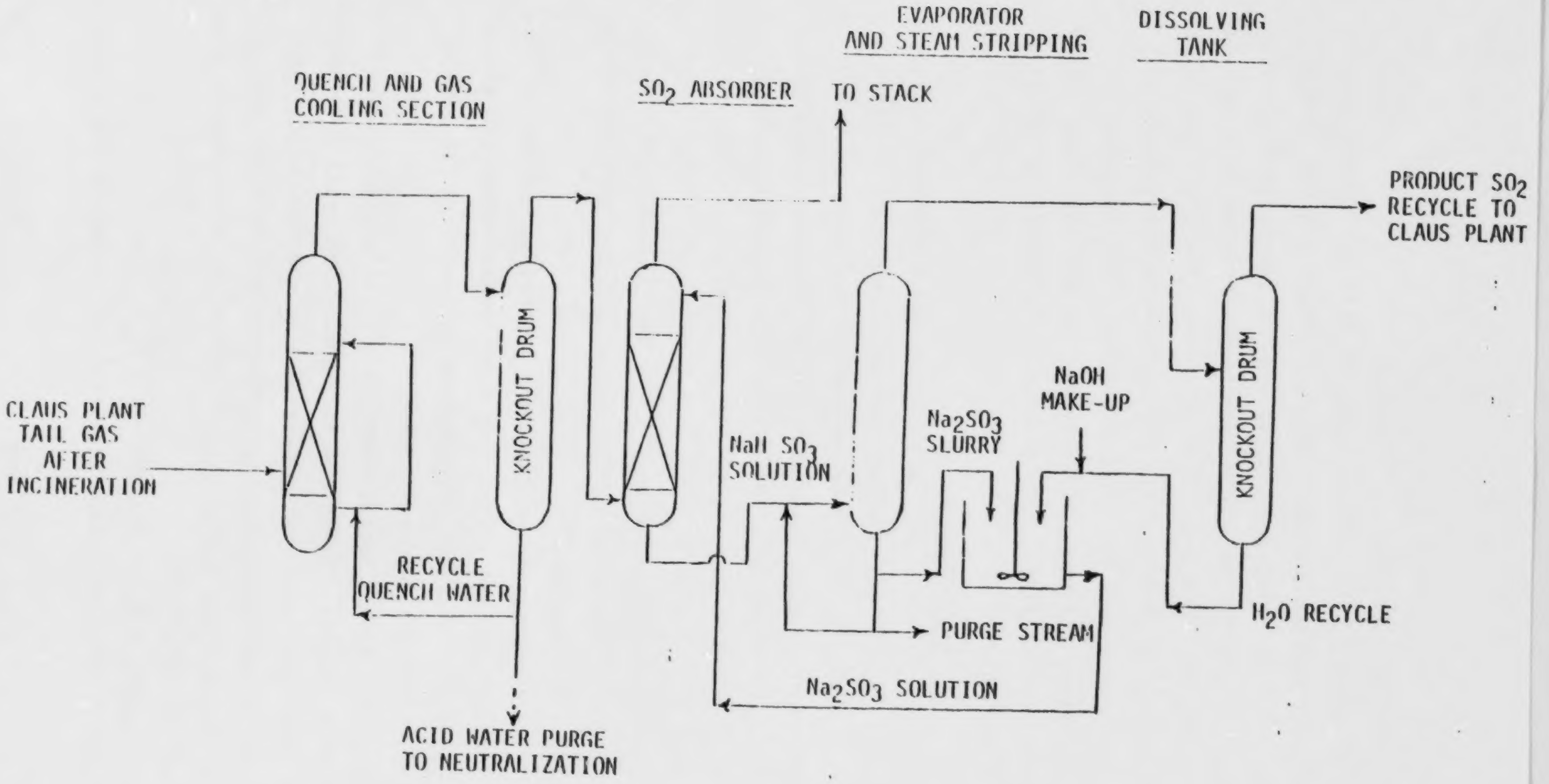

QUENCH AND GAS COOLING SECTION
SO_2 ABSORBER
TO STACK
EVAPORATOR AND STEAM STRIPPING
DISSOLVING TANK
PRODUCT SO_2 RECYCLE TO CLAUS PLANT
CLAUS PLANT TAIL GAS AFTER INCINERATION
KNOCKOUT DRUM
NaH SO_3 SOLUTION
Na_2SO_3 SLURRY
NaOH MAKE-UP
KNOCKOUT DRUM
RECYCLE QUENCH WATER
PURGE STREAM
H_2O RECYCLE
Na_2SO_3 SOLUTION
ACID WATER PURGE TO NEUTRALIZATION

Figure 4-2. Flow diagram for the Hellman-Lord SO_2 recovery process.

The bisulfite solution is boiled in an evaporator-crystallizer, where the bisulfite solution decomposes to SO_2 and H_2O vapor and sodium sulfite is precipitated:

$$2NaHSO_3 \xrightarrow{heat} Na_2SO_3\downarrow + H_2O + SO_2\uparrow$$

Sulfite crystals are separated and redissolved for reuse as lean solution to the absorber. The wet SO_2 gas is directed to a partial condenser where most water vapor is condensed and reused to dissolve sulfite crystals. The enriched SO_2 stream is then recycled back to the Claus plant for conversion to elemental sulfur or sent to an acid plant for conversion to sulfuric acid.

The Wellman-Lord process has been operating in U.S. refineries since 1972.

4.2.2 Reduction Tail Gas Scrubbers

At least four processes have been developed for tail gas sulfur removal. These processes convert the tail gas sulfur species to H_2S by a reduction step, then scrub the H_2S from tail gases prior to venting. These are the Beavon, Beavon MDEA, SCOT, and ARCO processes. The Beavon process is unique in that the H_2S is converted to sulfur outside the Claus unit using a lean H_2S-to-sulfur process called Stretford. The other three processes utilize conventional amine scrubbing and regeneration to remove the H_2S and recycle back as Claus feed. Since the Beavon MDEA, SCOT, and ARCO processes are similar and the SCOT process the most commonly used, the SCOT process will be described in more detail, with the Beavon MDEA and ARCO descriptions minimized to point out the deviations from the SCOT.

Also, since all processes utilize a reduction step, this step is described first as a common process.

4.2.2.1 The Reduction Step. All generic reduction tail gas processes utilize a reduction step in which sulfur species are converted essentially to H_2S by hydrogenation and hydrolysis under moderate conditions of temperature and pressure. Before the tail gas enters a packed bed hydrogenation reactor, fuel gas is combusted substoichiometrically in an inline burner to produce the reducing conditions necessary to convert sulfur gases to H_2S. The combustion products, primarily carbon monoxide (CO), nitrogen, and water vapor (H_2O), are mixed with the tail gas to provide a reducing atmosphere. Extra hydrogen may be required upstream of the burner, depending on the hydrogen content of Claus tail gas. A cobalt-molybdenum catalyst promotes the hydrogenation and hydrolysis reactions as follows:

$$S_8 + 8H_2 \rightarrow 8H_2S$$

$$SO_2 + 3H_2 \rightarrow H_2S + 2H_2O$$

$$COS + H_2O \rightarrow H_2S + CO_2$$

$$CS_2 + 2H_2O \rightarrow 2H_2S + CO_2$$

After hydrogenation and hydrolysis, the tail gas is cooled and water removed.

4.2.2.2 Beavon Process.[12,13] The Beavon process was developed by the Ralph M. Parsons Company and Union Oil Research.

In the Beavon or Beavon/Stretford process, the cooled gas is directed to [illegible] Stretford sulfur plant, where it is contacted countercurrently with a sodium solution and absorbed. The absorbed H_2S is oxidized and precipitated out of the solution as elemental sulfur solids, and the sodium values regenerated by the following reactions:

(a) Absorption of H_2S

$$H_2S + Na_2CO_3 \rightarrow NaHS + NaHCO_3$$

(b) Precipitation of sulfur

$$2NaVO_3 + NaHS + NaHCO_3 \rightarrow S\downarrow + Na_2V_2O_5 + Na_2CO_3 + H_2O$$

(c) Regeneration of sodium vanadate ($NaVO_3$)

$$Na_2V_2O_5 + ADA^* \text{ (oxidized)} \rightarrow 2NaVO_3 + ADA \text{ (reduced)}$$

* Anthraquinone Disulfonic Acid

Air is then blown through the solution to froth out the sulfur and regenerate the ADA:

$$ADA \text{ (reduced)} + 1/2\ O_2 \rightarrow ADA \text{ (oxidized)}$$

Sulfur froth is then collected, filtered, and remelted to be combined with Claus sulfur.

The overall reaction is the Claus reaction; hence, no chemicals are consumed in theory. Actually, adverse side reactions occur due to temperature excursions in the presence of trace oxidizing species in the tail gas, and result in the buildup of sodium thiosulfate and related compounds in the circulating liquor. This requires a periodic or continuous purge stream to keep dissolved solids to a desired level.

A new variation of the Beavon process involves replacement of the Stretford process with the Unisulf proces; although similar to the Stretford, the Unisulf reportedly requires no purge of solution under normal operating conditions.

Figure 4-3 is a typical flow diagram for the Beavon process.

Stretford absorber off-gases, typically containing 20-80 ppmv carbonyl sulfide and trace species of other sulfur gases, do not require incineration and are normally vented to the atmosphere without further processing. A stand-by incinerator is normally available, however, to handle process upsets where H_2S emissions exceed a given level, usually 10-20 ppmv in stack gases.

The Beavon process has been operating in U.S. refineries since 1973.

4.2.2.3 The SCOT Process.[14,15] The Shell Claus Off-gas Treating (SCOT) process scrubs the cooled reactor gas with an alkanolamine solution in an absorber. The solution selectively absorbs H_2S over SO_2. Absorbed acid gases are liberated from the amine solution by stripping with steam in a regenerator and are recycled to the gas inlet of the Claus unit.

Amine absorber off-gas containing about 200-300 ppmv H_2S requires incineration, but at a lower temperature (~540°C) than a typical Claus incinerator. A typical performance guarantee for the SCOT is 250 ppmv SO_2 in the incinerated off-gas, though guarantees as low as 150 ppmv have been given.

The SCOT process commonly uses diisopropanol amine (DIPA) a secondary amine or methyldiethanolamine (MDEA), a tertiary amine, which are more selective than amines used for refinery gas treating. Other solvents may be used, but the final choice depends on process economics.

Figure 4-4 schematically represents a typical SCOT process. The SCOT process has been operating in U.S. refineries since 1973.

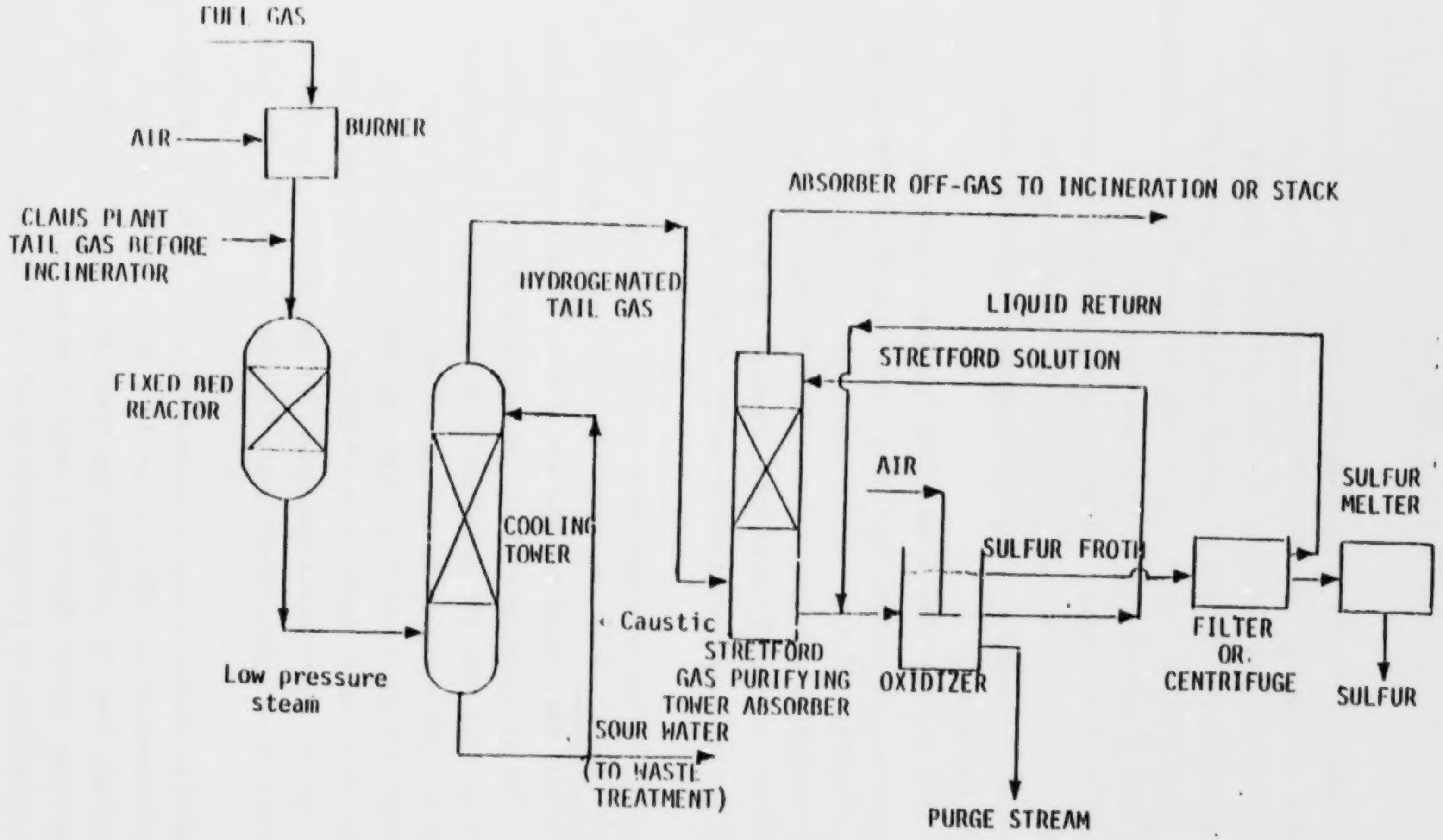

FUEL GAS
AIR
BURNER
CLAUS PLANT TAIL GAS BEFORE INCINERATOR
FIXED BED REACTOR
Low pressure steam
COOLING TOWER
HYDROGENATED TAIL GAS
Caustic
SOUR WATER (TO WASTE TREATMENT)
STRETFORD GAS PURIFYING TOWER ABSORBER
ABSORBER OFF-GAS TO INCINERATION OR STACK
LIQUID RETURN
STRETFORD SOLUTION
AIR
OXIDIZER
SULFUR FROTH
PURGE STREAM
FILTER OR CENTRIFUGE
SULFUR MELTER
SULFUR

Figure 4-3. Flow diagram for the Beavon sulfur removal process.

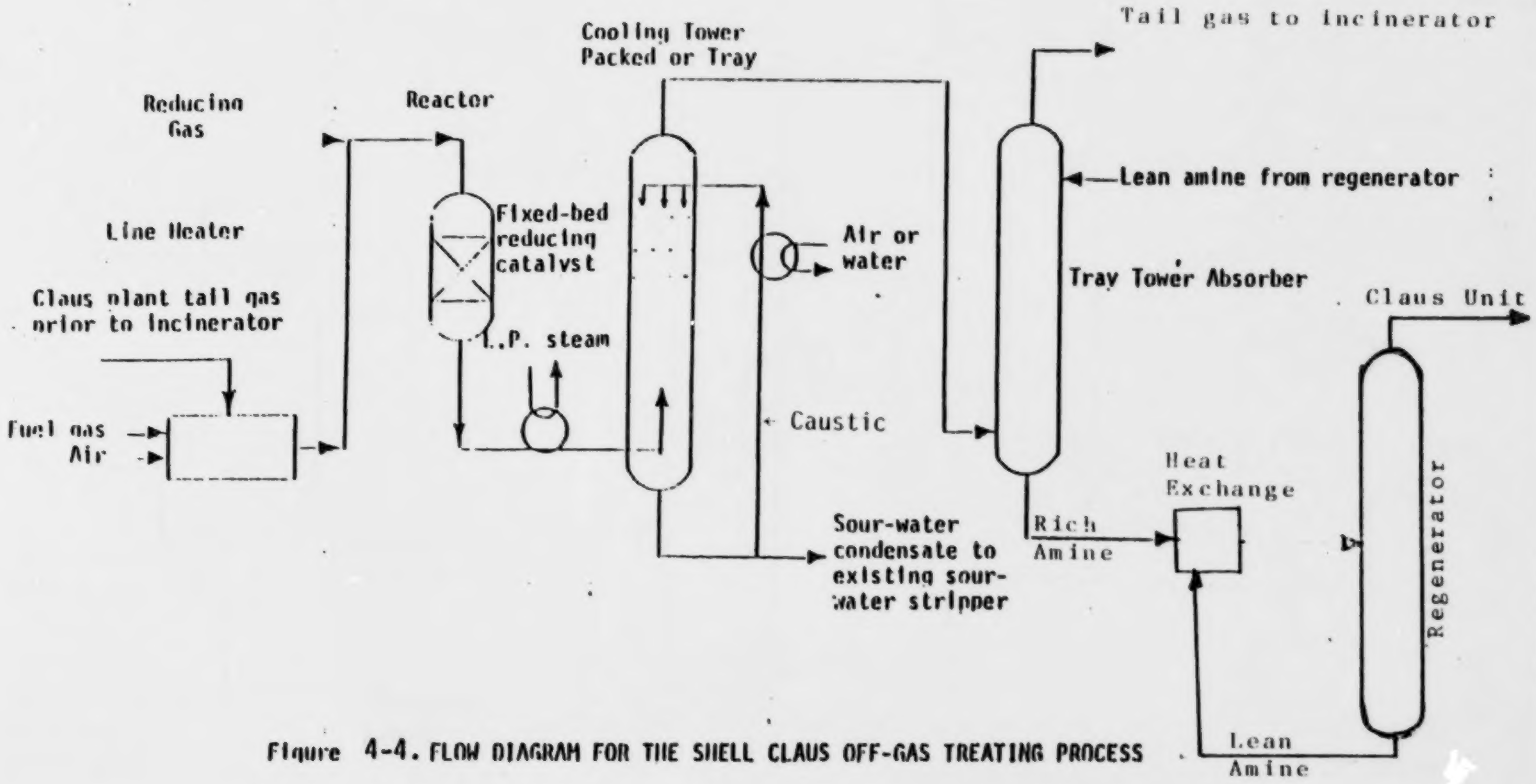
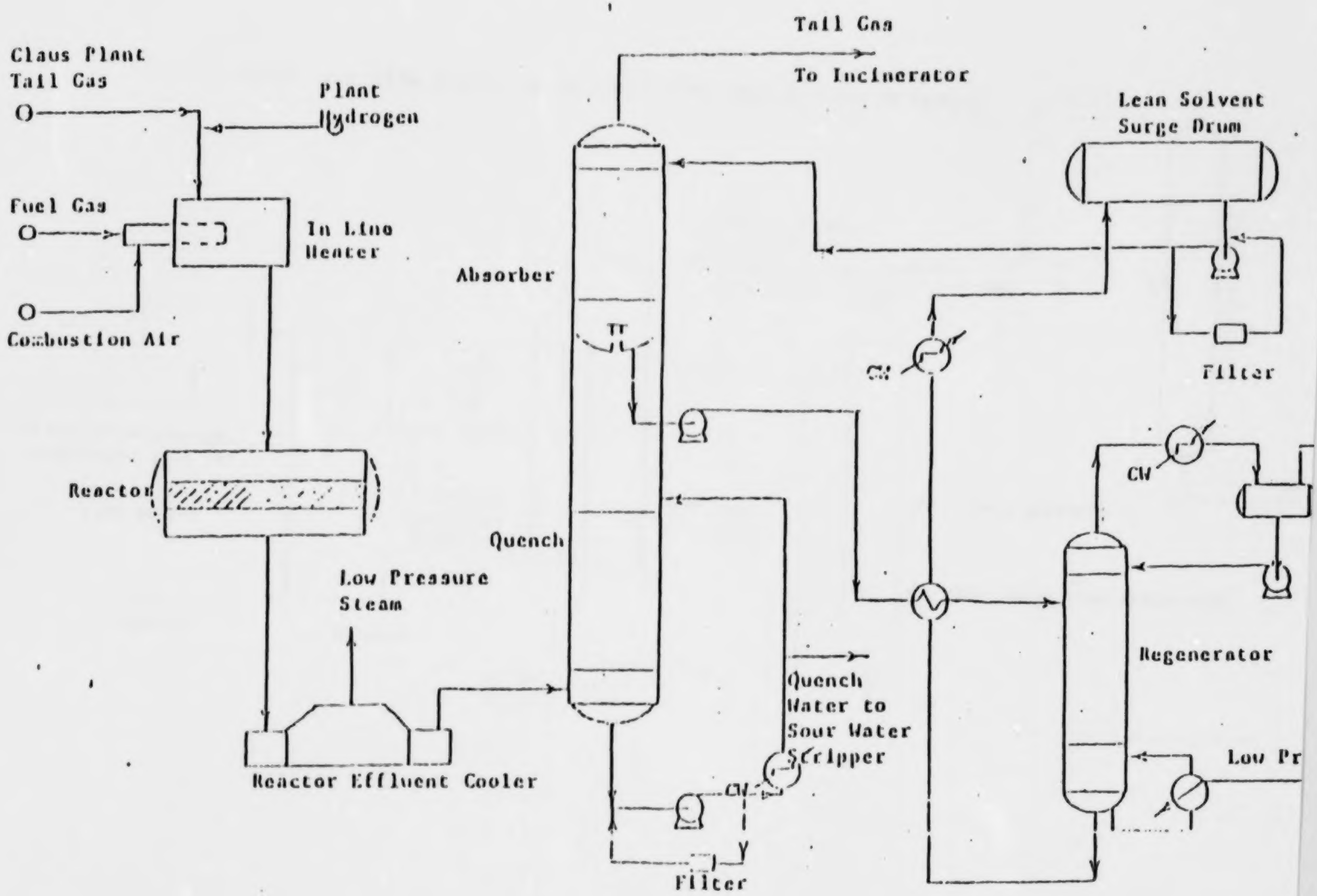

Reducing Gas
Line Heater
Claus plant tail gas prior to Incinerator
Fuel gas
Air
Reactor
Fixed-bed reducing catalyst
L.P. steam
Cooling Tower Packed or Tray
Air or water
Caustic
Sour-water condensate to existing sour-water stripper
Tail gas to Incinerator
Lean amine from regenerator
Tray Tower Absorber
Rich Amine
Heat Exchange
Claus Unit
Regenerator
Lean Amine

Figure 4-4. FLOW DIAGRAM FOR THE SHELL CLAUS OFF-GAS TREATING PROCESS

Figure 4-5. ARCO TAIL GAS PURIFICATION

Simplified Flow


Claus Plant Tail Gas
Plant Hydrogen
Fuel Gas
In Line Heater
Combustion Air
Reactor
Low Pressure Steam
Reactor Effluent Cooler
Absorber
Quench
Tail Gas
To Incinerator
Quench Water to Sour Water Stripper
Filter
CW
CW
Lean Solvent Surge Drum
Filter
CW
Regenerator
Low Pr

4.2.2.4 ARCO Process.[16] Conceptually similar to the SCOT process described above, the ARCO process is based upon amine absorption of H_2S and recycle to the Claus plant. Design performance levels of 250 ppmv SO_2 in the incinerated absorbed off-gas have been common for the ARCO process. Figure 4-5 is a representative ARCO process scheme. It has been installed in U.S. refineries since 1975.

4.2.2.5 Beavon/MDEA.[17] A recently announced option to the Beavon process previously described is substitution of the Stretford sulfur recovery plant with an amine absorber/regenerator with H_2S recycle to the Claus similar to the SCOT and ARCO processes. A representative schematic is not presented here, but it is assumed similar to the SCOT and ARCO processes, with associated performance guarantees. The Beavon/MDEA uses methyldiethanol amine (MDEA), a tertiary amine, which is more selective for H_2S than the secondary amines frequently used in amine tail gas processes. Also, the licensors prefer to generate all needed hydrogen in the reducing gas generator, obviating an external source of hydrogen.

4.3 COMMERCIAL STATUS OF EMISSION CONTROLS FOR REFINERY SULFUR PLANTS[18,19,20,21,22]

The first commercial tail gas treater installed in 1972 in a U.S. refinery was the Wellman-Lord process. The Beavon, SCOT, and IFP-1 processes were installed at U.S. refineries the following year. In 1975, the first ARCO process was installed. Since 1976, when the NSPS for refinery sulfur plants was announced, all sulfur plants subject to the NSPS have chosen the SCOT, Beavon, or the ARCO processes, although one non-NSPS Wellman-Lord unit was installed in 1981. Table 4-1 lists the tail gas units installed in U.S. refineries as of 1982. Units planned or under construction are listed in Table 4-2. Each "unit" refers to a separate tail gas process sequence as shown in Figures 4-1 through 4-5. A unit may serve one or several Claus units. Capacities shown in Table 4-1 are for total Claus capacity served.

Table 4-1. TAIL GAS TREATERS INSTALLED IN U.S. REFINERIES

Unit	Location	Onstream Date	No. of Units	Total Sulfur Plant Capacity, Mg/D (LT/D)
ARCO	California	1975	1	182.9 (180)
ARCO	Texas	1976	1	320.1 (315)
ARCO	Washington	1977	1	122.0 (120)
ARCO	Pennsylvania	1982	1	172.8 (170)
Beavon	California	1973	2	203.2 (200)
Beavon	California	1973	2	304.8 (300)
Beavon	Pennsylvania	1973	1	142.3 (140)
Beavon	California	1974	1	355.7 (350)
Beavon	California	1975	3	249.0 (245)
Beavon	Louisiana	1975	3	312.0 (307)
Beavon	Louisiana	1976	1	304.8 (300)
Beavon	Louisiana	1976	1	235.8 (232)
Beavon	New Jersey	1976	1	304.8 (300)
Beavon	Texas	1977	1	304.8 (300)
Beavon	Texas	1977	1	829.3 (816)
Beavon	Illinois	1977	2	304.8 (300)
Beavon	Louisiana	1978	1	235.8 (232)
Beavon	New Jersey	1980	2	274.4 (270)
Beavon	Texas	1980	1	101.6 (100)
Beavon	Missouri	1981	1	233.8 (230)
Beavon	Indiana	1981	1	396.4 (390)

Table 4-1. (Continued)

Unit	Location	Onstream Date	No. of Units	Total Sulfur Plant Capacity, Mg/D (LT/D)
Beavon	California	1981	2	122.0 (120)
Beavon	California	1981	1	152.5 (150)
Beavon	Louisiana	1982	1	203.2 (200)
Beavon	California	1982	1	39.6 (39)
IFP-1	Texas	1973	1	45.7 (45)
IFP-1	Texas	1973	1	45.7 (45)
IFP-1	Texas	1976	1	101.6 (100)
IFP-1	California	1976	1	182.9 (180)
IFP-1	Texas	1976	1	406.4 (400)
IFP-1	Texas	1977	1	254.1 (250)
SCOT	California	1973	1	15.2 (15)
SCOT	California	1973	1	35.6 (35)
SCOT	Pennsylvania	1974	1	162.6 (160)
SCOT	Michigan	1975	1	81.3 (80)
SCOT	Oklahoma	1975	1	29.5 (29)
SCOT	Louisiana	1975	1	42.7 (42)
SCOT	Texas	1975	1	318.1 (313)
SCOT	Louisiana	1976	1	43.7 (43)
SCOT	Texas	1976	1	196.2 (193)
SCOT	Texas	1977	1	152.5 (150)
SCOT	Oklahoma	1977	1	63.0 (62)

Table 4-1. (Continued)

Unit	Location	Onstream Date	No. of Units	Total Sulfur Plant Capacity, Mg/D (LT/D)
SCOT	Texas	1977	1	233.8 (230)
SCOT	Pennsylvania	1978	1	46.8 (46)
SCOT	Louisiana	1979	1	152.5 (150)
SCOT	California	1979	1	10.2 (10)
SCOT	Illinois	1979	1	457.4 (450)
SCOT	Wyoming	1980	1	50.8 (50)
SCOT	Texas	1980	1	381.1 (375)
SCOT	Ohio	1980	1	122.0 (120)
SCOT	Ohio	1980	1	101.6 (100)
SCOT	California	1980	1	7.4 (7.3)
SCOT	California	1981	1	177.9 (175)
SCOT	Kentucky	1981	1	203.2 (200)
SCOT	Texas	1982	1	115.9 (114)
SCOT	Louisiana	1982	1	127.0 (125)
SCOT	Louisiana	1982	1	61.0 (60)
SCOT	Louisiana	1982	1	8.1 (8)
SCOT	Texas	1982	1	14.2 (14)
SCOT	Alabama	1982	1	40.7 (40)
SCOT	Texas	1982	1	18.3 (18)
SCOT	Alabama	1982	1	53.9 (53)

Table 4-1. (Continued)

Unit	Location	Onstream Date	No. of Units	Total Sulfur Plant Capacity, Mg/D (LT/D)
Wellman-Lord	California	1972	1	457.4 (450)
Wellman-Lord	California	1975	1	330.3 (325)
Wellman-Lord	California	1976	1	304.8 (300)
Wellman-Lord	California	1977	1	330.3 (325)
Wellman-Lord	California	1981	1	203.2 (200)

Table 4-2. REFINERY TAIL GAS UNITS PLANNED OR UNDER CONSTRUCTION

Unit	Location	Onstream Date	No. of Units	Total Sulfur Plant Capacity, Mg/D (LT/D)
Beavon/MDEA	Louisiana	1983	2	365.9 (360)
Beavon/MDEA	Louisiana	1983	2	233.8 (230)
Beavon	Alaska	--	1	229.7 (226)
Beavon	Kansas	--	1	10.5 (10.3)
Beavon	California	--	1	30.5 (30)
SCOT	Tennessee	1983	1	91.5 (90)
SCOT	Texas	1983	1	4.6 (4.5)
SCOT	California	1983	1	304.8 (300)
SCOT	Texas	1983	1	252.1 (248)
SCOT	Texas	1983	1	255.1 (251)
SCOT	Minnesota	1983	1	304.8 (300)
SCOT	Washington	1983	1	50.8 (50)
SCOT	Texas	1983	1	79.3 (78)
SCOT	California	1983	1	66.1 (65)
SCOT	Delaware	1984	1	241.7 (235)
SCOT	Louisiana	1984	1	132.1 (130)
SCOT	Texas	1985	2	1,016.4 (1,000)
SCOT	Louisiana	--	1	38.6 (38)
SCOT	Ohio	--	1	32.5 (32)
SCOT	Texas	--	1	162.6 (160)
SCOT	Texas	--	1	177.9 (175)

As shown in these tables, there are 76 reported tail gas treaters operating in domestic refineries with an additional 24 units planned or under construction. These figures do not account for units that have been replaced or are currently inoperative.* Total sulfur plant capacity controlled by these units is 12,514 Mg/D with an additional 4,109 Mg/D planned or under construction. Thus, the average tail gas treater currently operating handles 165 Mg/D of Claus plant capacity, while planned units average 179 Mg/D Claus capacity.

*Also not included is a hybrid 34.9 Mg/D tail gas unit which is not commercially available.

4.4 REFERENCES

1. Kerr, et al. "A new sulfur recovery process, The RSRP", Oil & Gas Journal, July 26, 1982, pp. 230-243.

2. "Sulfur Recovery Study--Onshore Sour Gas Production Facilities", Ralph M. Parsons Company. July 1981. p. 4-9.

3. Letter, J. C. Brocoff, Ralph M. Parsons Company, to S.T. Cuffe, U.S. EPA, dated February 16, 1983.

4. Letter, J. B. Lartigue, Aquitaine Company of Canada, Inc., to James Durham, U.S. EPA. January 14, 1975.

5. "SO_2 Emissions in Natural Gas Production Industry - Background Information For Proposed Standards", EPA 450/3-82-023a, January 1983.

6. Goddin, C. S. etal, "CBA Process vs. Claus Recovery" - Hydrocarbon Processing. October 1974. pp. 122-124.

7. Letter, D. H. Dilworth, Davy McKee Corporation, to C. Sedman, U.S. EPA, January 12, 1983.

8. Hirai, M. etal, "Solvent/Catalyst Mixture Desulfurize Claus Tail Gas", Chemical Engineering. April 17, 1972.

9. Trip Report - "Visit to IFP Sulfur Recovery Unit", C. Sedman, U.S. EPA. April 18, 1974.

10. Craig, T. L., "Tail-gas Desulfurization Operations Successful", Oil & Gas Journal. February 7, 1982. pp. 65-67.

11. Reference 3. p. 4-25.

12. Reference 2. p. 4-5.

13. Reference 3.

14. Reference 2. pp. 4-14 and 4-15.

15. Kuijpers, N.S.M.J., "The Shell Off-Gas Treating Process" - presented at the Gas Sweetening and Sulfur Recovery Seminar, Amsterdam, The Netherlands. November 9-13, 1981.

16. Trip Report - "Visit to ARCO Refinery, Pasadena, Texas", Charles Sedman, U.S. EPA. September 20, 1982.

17. Reference 2. p. 4-3

18. Letter, W. T. Knowles, Shell Oil Company, to Charles B. Sedman, U.S. EPA. August 24, 1982.

19. Letter, M. A. Peterson, Union Oil Company of California, to Charles B. Sedman, U.S. EPA. September 15, 1982.

20. "Survey Report on SO_2 Control Systems for Non-Utility Combustions and Process Sources - May 1977", prepared by PEDCo Environmental, Inc. Contract No. 68-02-2603.

21. Letter, H. J. Grimes, ARCO Petroleum Products Co. to C. Sedman, U.S. EPA, dated October 5, 1982.

22. Letter, D. H. Dilworth, Davy McKee, to C. Sedman, U.S. EPA, dated October 5, 1982.

5. COMPLIANCE STATUS OF REFINERY SULFUR PLANTS

5.1 AFFECTED FACILITIES

Of the 43 sulfur plants constructed during the period 1977-1982 in domestic petroleum refineries, only 17 are subject to the sulfur plant NSPS. Of the 26 non-NSPS units, only 4 were exempted due to size (less than 20 long tons per day capacity). The remaining 22 units were contracted for prior to October 4, 1976, and were "grandfathered" as an existing facility at the time of NSPS proposal.

Of the 17 units subject to the NSPS, 7 are in start-up and have not been compliance tested. Emission test results from the 10 certified NSPS facilities are presented and discussed in the following section. Unless otherwise noted, all results are based on three test runs using EPA methods discussed in Chapter 3.

5.2 COMPLIANCE TEST RESULTS

5.2.1 Reduced Sulfur and Hydrogen Sulfide[1,2,3,4]

As discussed in Chapter 3, reduced sulfur compounds and hydrogen sulfide limits are enforced wherever a reduction tail gas system is used and the tail gas not incinerated after treatment. Four Beavon tail gas units are operating under these restrictions, and the compliance test results are summarized in Table 5.1.

Table 5.1 illustrates the effectiveness of the Beavon process, especially the Stretford H_2S absorber. Of the four units tested, all are in compliance, being well under the 300 ppmv reduced sulfur and 10 ppmv H_2S restrictions. Typically, the only measurable sulfur compound present in Beavon exhaust gases is carbonyl sulfide (COS).

5.2.2 Sulfur Dioxide[5,6,7,8,9,10]

Units which incinerate tail gases are subject to sulfur dioxide limits of 250 ppmv dry basis, corrected to zero percent oxygen. Six SCOT treaters which incinerate tail gas after treatment are currently operating under these rules, and the associated emission test results are presented in Table 5-2.

Table 5-1. NSPS COMPLIANCE TEST RESULTS FOR REDUCED SULFUR & HYDROGEN SULFIDE

Plant	Size, Mg/D*	Average Emissions, ppmv** Reduced Sulfur	H_2S
A	101.6	16	4.3
B	233.8	2	<1
C	235.8	62	8.5
D	235.8	161	<4

* Parent Claus Capacity
** EPA Method 15

Table 5-2. NSPS COMPLIANCE TEST RESULTS FOR SULFUR DIOXIDE

Plant	Size, Mg/D*	Average SO_2 Emissions, ppmv
E	122.0	168**
F	381.1	112***
G	101.6	205
H	203.2	183
I	61.0	203
J	40.7	81

* Parent Claus capacity
** Average of 20 runs from four separate tests
*** State test method; all others EPA-6

As shown in Table 5-2, the SCOT emissions are somewhat higher than for Beavon units, and somewhat less predictable, reflecting the effect of process conditions upon the amine absorbers. Of the six units tested, average emissions range from approximately 100 to 200 ppmv SO_2. All six units are in compliance.

These short-term tests represent the only emission data gathered during this study. Although SO_2 and reduced sulfur monitors are generally installed on these NSPS units, data are not recorded and reported to agencies and are, therefore, not available for analysis.

5.3 OPERABILITY OF NSPS UNITS[10,11,12,13,14,15,16]

Through EPA and API surveys, a total of 7 NSPS and 16 non-NSPS refineries responded to questions concerning operability and maintenance problems encountered in tail gas treaters.

From the surveys, it is evident that most problems in tail gas treaters are preceded by upsets in the Claus plant, which can send excessive amounts of either SO_2 or H_2S into the tail gas reactor. For an amine tail gas system, unchecked breakthrough of SO_2 through the reactor into the absorber causes no immediate excess emission because the amine combines irreversibly with SO_2. However, permanent loss of solution activity ensues, the solution becomes corrosive, and requires discarding. A breakthrough of H_2S beyond the design capacity of the absorber causes excess emissions of H_2S, but solution performance returns to normal as soon as the breakthrough is stopped.

A short-term breakthrough of SO_2 into the Stretford system causes no excess emissions because the Stretford solution also reacts irreversibly with SO_2 causing an increase in chemical consumption and more frequent system purge. The same is true of short-term H_2S overloads above design capacity, but prolonged overloads cause tower plugging and adversely affect Stretford chemicals which may take several days to return to normal operation.

The above helps to explain the survey results which show:

- ° older, non-NSPS units to be more reliable than NSPS units (increased reliability with system age)
- ° most problems directly attributable to SO_2 breakthrough

Common problems reported for amine systems included excess solvent foaming, quench water filter plugging, quench column level control, and catalyst bed plugging. Less frequent problems included heater tube leaks, pump failures, and blower failures, all of which appear unrelated to the process itself.

Similar reactor and quench tower problems were reported for Stretford units, along with the less routine pump, compressor, and heat exchanger failures. Additionally, the Stretford portion of some units using direct melting of sulfur slurry has caused less severe, but more consistent, maintenance problems. Plugging of decanter and melters along with general solids accumulation have been reported.

Generally, the survey indicates the most important factor in successful tail gas plant operation is experience. For units with more than 3 years operating experience (mostly non-NSPS units), system reliabilities approach 100 percent in many cases. Both amine and Stretford units received praise from operators. However, the vast majority of problems and somewhat less enthusiastic responses to the survey came from NSPS units.

Most SO_2 and H_2S breakthrough-related problems (quench tower plugging and corrosion, high chemical consumption) appear corrected by closer attention to the built-in safeguards in tail gas treaters. The alkaline guard (quench tower pH control) and level control should alleviate most downstream corrosion, plugging, and chemical degradation problems. Operating at a $H_2S:SO_2$ ratio slightly above 2 to 1 allows for a greater margin of operating error without irreversible loss of solution activity or onset of corrosion problems.

Reactor problems appear due to the introduction of unsaturated hydrocarbons via fuel gas to the heater and should be alleviated by better quality control of fuel.

Degradation of amines and excess foaming have been alleviated by installation of carbon absorption units and use of anti-foaming agents.

Stretford problems involving plugging and solids accumulation have been alleviated by replacement of level controllers and more operator attention. Stretford solutions outfitted to filter and rinse sulfur before melting have been more successful, and the licensor is exclusively using filters in new plants under design.[17]

5.4 STATUS OF EMISSION MONITORS[18]

5.4.1 SO_2 Monitors

Where incinerators are used to oxidize tail gas, sulfur dioxide monitors have been installed on all new units surveyed. Practically all existing tail gas installations with incinerators also use SO_2 monitors. Both in-stack and extractive type SO_2 monitors, identical to those found on boilers, are currently operating. Problems encountered are similar to those on boilers, and include:

- plugged sampling lines on extractive systems
- probe failures on extractive systems
- sample conditioning system on in-stack monitors
- factory servicing of in-stack monitors

Most in-stack monitors installed prior to 1980 performed very poorly in field applications and required reservicing at the factory or replacement with more durable instrumentation. Vendors have also made improvements in sample extraction and conditioning components, as evidenced by the improved reliabilities reported by more recent installations.

Extractive monitors have experienced initial problems with the sampling lines and probes. Installation of probe shields and higher pressure backflush systems in sample lines have alleviated these problems.

5.4.2 Reduced Sulfur and H_2S Monitors

Reduced sulfur monitors are relatively new and were found on only two operating facilities. In both cases, the systems were reported as unsatisfactory due to high maintenance and poor operability. Problems encountered include probe and sample line plugging, and several failures of the computer software which required reprogramming.

Hydrogen sulfide monitors are generally the lead acetate tape monitors which are used in conjunction with an H_2S alarm system tied to a standby incinerator. As such, these monitors are more qualitative than quantitative and would not meet stringent performance criteria. Problems reported are minimal and often were due to lack of periodic maintenance.

5.5 EMISSION TESTING

One small consideration should be noted with regard to EPA Method 15--determination of reduced sulfur compounds. Most recent emission tests have been performed using a modified EPA Method 15, where acetate buffer and improved chromatographic separation columns have simplified the sample conditioning requirements of Method 15.[19]

5.6 REFERENCES

1. Letter, R. T. Denbo, Exxon Company, U.S.A., to Don R. Goodwin, U.S. EPA, dated June 11, 1982.

2. Letter, G. E. Lowe, Marathon Petroleum Company, to Don R. Goodwin, U.S. EPA, dated September 17, 1982.

3. Letter, R. J. Niederstadt, Mobil Oil Corporation, to Don R. Goodwin, U.S. EPA, dated June 15, 1982.

4. Letter, Steven Feeler, Missouri Department of Natural Resources, to C. B. Sedman, U.S. EPA, dated September 24, 1982.

5. Letter, C. M. Tyler, SOHIO, to Don R. Goodwin, U.S. EPA, dated July 15, 1982.

6. Letter, J. P. Gay, Ashland Petroleum, to Charles B. Sedman, U.S. EPA, dated September 27, 1982.

7. Letter, B. F. Ballard, Phillips Petroleum, to Don R. Goodwin, U.S. EPA, dated July 13, 1982.

8. Letter, Richard Grusnick, Alabama Department of Environmental Management, to Charles B. Sedman, U.S. EPA, dated October 15, 1982.

9. Letter from G.J. Vetter, GHR Energy Corporaiton, to C. Sedman, U.S. EPA, dated January 28, 1983.

10. Letters, E. P. Crockett, American Petroleum Institute, to Charles B. Sedman, U.S. EPA, dated June 15, June 30, and July 14, 1982.

11. Reference 1.

12. Reference 2.

13. Reference 3.

14. Reference 6.

15. Reference 8.

16. Letter, L. M. Lovell, Amoco Oil Company, to Don R. Goodwin, U.S. EPA, dated June 23, 1982.

17. Letter from J. C. Brocoff, Ralph M. Parsons Co., to S. T. Cuffe, U.S. EPA, February 16, 1983.

18. References 11-17.

19. Telephone Conversation, B. Ferguson, Harmon Engineering and Testing, Inc., to C. Sedman, U.S. EPA, dated November 18, 1982.

6. MODEL PLANTS AND COST ANALYSES

This chapter defines model plants which represent typical refinery sulfur plant alternatives for new installations and presents estimated costs of those alternatives.

6.1 MODEL PLANTS

In order to have a common basis for comparing costs of emission controls to meet the existing NSPS, model plants are selected. Resource requirements, dollar costs, and environmental impacts are then determined for each model plant. From these assessments, the relative impact and appropriateness of NSPS for various size sulfur plants may be weighed.

6.1.1 Model Plant Size

In Chapter 4 it was shown that sulfur plants constructed with tail gas treaters since 1972 have ranged from 7.4 to 457.4 megagrams per day (Mg/D) capacity. Actual individual sulfur plants up to 400 Mg/D have been constructed. Planned tail gas units range from 4.6 to 1,016 Mg/D, with single Claus plants of up to 508 Mg/D forecasted. Tail gas units constructed in the United States have been either the extended Claus systems (IFP) or add-on absorbers (Wellman-Lord, SCOT, Beavon, or ARCO). All planned tail gas units are essentially the reduction/absorption type, with the amine scrubbing variation representing the majority choice.

For the economic modelling and comparisons, Claus plants at 10.16, 50.8, and 101.6 Mg/D have been selected for model analyses.

6.1.2 Choice of Representative Control System

The NSPS control cases are represented by the reduction/amine absorption process for simplicity. Although the oxidation (Wellman-Lord) system is clearly an alternative, the reduction systems have been the overwhelming choice for NSPS Claus plants. The Beavon-Stretford process has certain advantages over the amine (SCOT/ARCO/Beavon-MDEA) systems with respect to increased size and decreased H_2S content in Claus feed; however, for typical refinery applications in the 10 to 100 Mg/D range, amine systems are the majority (18 of 20 operating units) choice for new installations (see Appendix A, pg. A-8 for more discussion).

6.1.3 Assumptions of Modelling Parameters

Table 6-1 presents process parameters of model plants chosen. These model plants were developed using reported process data from NSPS plants, technical data from vendors, and previous studies of sulfur recovery plants by EPA.[1,2] Details of each model are discussed in Appendix A.

All cases handle acid gas consisting of 80 percent hydrogen sulfide, 10 percent carbon dioxide, 4.5 percent ammonia, 0.5 percent hydrocarbons, and 5.0 percent moisture. The acid gas streams are assumed saturated at 42.9°C (109°F) and 170 kilopascals (24.7 psia). Sour water streams containing the bulk of hydrocarbons and all ammonia are completely combusted in the first combustion stage, with amine off-gases combusted in the second stage.

Claus plants are assumed to use high efficiency alumina catalysts for maximum sulfur recovery: the 101.6 LT/D case uses two Claus stages at 95.1 percent recovery, while the 50.8 and 101.6 Mg/D cases use three Claus stages at 96.6 percent recovery.

Tail gas units are sized at twice the anticipated feed rate, and Claus plants are sized to accomodate the additional recycle stream. For example, the model plant 3B features a 105.0 Mg/D Claus plant (101.6 Mg/D feed, 3.4 Mg/D recycle, 0.1 Mg/D emission rate) and a tail gas unit sized at 6.8 Mg/D. Since the recycle stream is more dilute with respect to H_2S, the Claus size (based on gas flow) actually increases by 50 percent in the 3-stage cases and 7.6 percent in the 2-stage case.

All Claus plants consume 4,300 Kp steam and generate 1,760 Kp and 106 Kp steam, with 3-stage plants also generating 352 Kp steam. Boiler feedwater is available at 2,255 Kp and 110°C, while cooling water is available at 29°C and returned at 43°C. Incinerators are designed to operate at 649°C (1200°F), 25 percent excess air for the Claus only cases, and the Claus/tail gas/incinerator heat recovery case. Incinerators operate at 538°C (1000°F), 25 percent excess air for tail gas treating with no incinerator heat recovery. Only for the 101.6 Mg/D case is waste heat recovery employed at the incinerator.

Table 6.1. MODEL PLANT PARAMETERS

MODEL PLANT 1A

1. Sulfur Intake: 10.16 Mg/D (10 LT/D)
2. Sulfur recovered: 9.66 Mg/D (95.10% efficiency)
3. Plant description: Claus furnace, two catalytic stages + incinerator
4. SO_2 emission rate: 348.6 Mg/Y (384.2 T/Y)
5. Operating schedule: 350 D/Y

MODEL PLANT 1B

1. Sulfur intake: 10.16 Mg/D (10 LT/D) + 0.49 Mg/D recycle
2. Sulfur recovered: 10.15 Mg/D (99.90 percent efficiency)
3. Plant description: Claus furnace, two catalytic stages, one catalytic reactor, amine absorption and regeneration, incinerator
4. SO_2 emission rate: 7.1 Mg/Y (7.84 T/Y)
5. Operating schedule: 350 D/Y

MODEL PLANT 2A

1. Sulfur intake: 50.80 Mg/D (50 LT/D)
2. Sulfur recovered: 49.09 Mg/D (96.64% efficiency)
3. Plant description: Claus furnace, three catalytic stages + incinerator
4. SO_2 emission rate: 1,209.4 Mg/Y (1,332.8 T/Y)
5. Operating schedule: 350 D/Y

MODEL PLANT 2B

1. Sulfur intake: 50.80 Mg/D (50 LT/D) + 1.68 Mg/D recycle
2. Sulfur recovered: 50.75 Mg/D (99.90% efficiency)
3. Plant description: Claus furnace, three catalytic stages, one catalytic reactor, amine absorption and regeneration, incinerator

Table 6.1. MODEL PLANT PARAMETERS (continued)

4. SO_2 emission rate: 35.56 Mg/Y (39.20 T/Y)

5. Operating schedule: 350 D/Y

MODEL PLANT 3A

1. Sulfur intake: 101.6 Mg/D (100 LT/D)

2. Sulfur recovered: 98.15 Mg/D (99.64 percent efficiency)

3. Plant description: Claus furnace, three catalytic stages, incinerator with heat recovery

4. SO_2 emission rate: 2,418.9 Mg/Y (2,665.6 T/Y)

5. Operating schedule: 350 D/Y

MODEL PLANT 3B

1. Sulfur intake: 101.6 Mg/D (100 LT/D) + 3.35 Mg/D recycle

2. Sulfur recovered: 101.5 Mg/D (99.90% efficiency)

3. Plant description: Claus furnace, three catalytic stages, one catalytic reactor, amine absorption and regeneration, incinerator with heat recovery

4. SO_2 emission rate: 71.12 Mg/Y (78.40 T/Y)

5. Operating schedule: 350 D/Y

With waste heat recovery, 600 psig steam is also generated, while tail gas treaters are net consumers of 50 psig steam. Complete utility consumption and generation balance sheets are presented in Appendix A to this document.

6.2 CONTROL LEVELS

Basically, the control levels are represented by the two sulfur recovery levels currently achieved in actual practice--96.6 percent recovery or control for the basic 3-stage Claus with alumina catalysts and 99.9 percent recovery for 3-stage Claus with state-of-the-art tail gas controls represented by amine absorption/recycle processes. (For 2-stage smaller sulfur plants, 95.1 percent recovery is achieved with a proportionally larger tail gas system to achieve 99.9 percent overall control.) Henceforth the Claus-only case will be referred to as baseline control and the Claus and tail gas treatment as NSPS control.

6.3 COST ANALYSIS

The model plants described in Section 6.1 were analyzed for economic impacts of controls by estimating fixed capital costs, annualized costs, emission reductions, and cost-effectiveness of controls. The estimates are based upon previous sulfur plant studies and the data from actual new installations as gathered by EPA specifically for this study. Detailed cost analyses are presented and discussed in Appendix A to this report.

6.3.1 Assumptions

Fixed capital costs were estimated from an analysis of capital cost data furnished by individual operating plants and equipment vendors. The range of operating variables examined were so great that a composite model facility was selected with distinct modelling and economic assumptions. Modelling assumptions were presented in Table 6.1. Table 6.2 lists key economic assumptions used to determine representative annualized costs. The most difficult economic parameter to gauge is the assignment of maintenance and repair costs. Previous studies have used vendor projections of maintenance costs at 3.5% of fixed capital costs;[1,2] while the background document to the original NSPS estimated maintenance costs at 3 percent of fixed capital for tail gas treaters.[3]

Table 6-2. ECONOMIC ASSUMPTIONS USED TO CALCULATE ANNUALIZED COSTS[a]

I. Utility prices:

1.	4,300 Kp steam	$\$15.98$/Mg	($\$7.25$/1,000 lb)
2.	1,760 Kp steam	$\$14.88$/Mg	($\$6.75$/1,000 lb)
3.	352 Kp steam	$\$12.68$/Mg	($\$5.75$/1,000 lb)
4.	106 Kp steam	$\$ 9.92$/Mg	($\$4.50$/1,000 lb)
5.	boiler feedwater	$\$ 3.31$/Mg	($\$1.50$/1,000 lb)
6.	steam condensate	$\$ 2.76$/Mg	($\$1.25$/1,000 lb)
7.	cooling water	$\$13.21/10^3\text{m}^3$	($\$$ .05/1,000 gal)
8.	catalyst:		
	a. alumina	$\$500/\text{m}^3$	($\$17/\text{ft}^3$)[b]
	b. cobalt-molybdenum (Co/Mo)	$\$5,000/\text{m}^3$	($\$170/\text{ft}^3$)[b]
9.	Chemicals:		
	a. diisopropanolamine	$\$0.49$/Kg	($\$1.07$/lb)[c]
	b. soda	$\$330.6$/Mg	($\$300$/ton)[c]
10.	fuel gas	$\$3.64/10^9$/J	($\$3.50/10^6$ Btu)[d]
11.	electric power	$\$0.05$/KWH	
12.	sulfur	$\$98.42$/Mg	($\$125$/LT)[e]

II. Labor (8,720 hours per year basis)

1. operators: ($14.50/hr)
   2/3 per shift for Claus
   2/3 per shift for tail gas treater

2. supervision: ($18.80/hr)
   1/4 per shift for sulfur recovery facility

III. Maintenance and Repair

Labor and materials: 3.0 percent of fixed capital Costs[c]

IV. Other Miscellaneous Costs

1. Operating supplies: 10 percent of operating labor
2. Laboratory charges: 10 percent of operating labor

V. Fixed Charges

1. Capital charges = fixed capital costs x $\frac{i(1+i)^n}{(1+i)^n-1}$
   = .13147 for n = 15 years, i = 10 percent
   = .171059 for n = 15 years, i = 15 percent
   = .213821 for n = 15 years, i = 20 percent

2. Local taxes - 1 percent of fixed capital costs

3. Insurance - 0.6 percent of fixed capital costs

Table 6.2. ECONOMIC ASSUMPTIONS USED TO CALCULATE ANNUALIZED COSTS[a] (continued)

VI. Overhead

1. plant overhead - 25 percent of operating labor + 25 percent of maintenance and repair
2. administrative - 1 percent of annualized costs
3. distribution and marketing - 1 percent of annualized costs

---

[a] All assumptions and values assigned from Reference 1 unless otherwise noted; actual consumption figures for model plants from EPA survey and Reference 2.

[b] Telephone conversation with Mr. R. E. Warner of Ralph M. Parsons Co., Feb. 1, 1983.

[c] Chemical Market Reporter, October 4, 1982.

[d] Memorandum: R. E. Jenkins to C. B. Sedman, EPA, dated September 7, 1982.

[e] Average of EPA survey.

Actual maintenance costs gathered by EPA for this study showed Claus costs ranging from 2.3 percent to 6.1 percent of fixed capital costs and 2.1 to 6.3 percent of fixed capital for reduction-based tail gas units. Estimates chosen for this study estimated maintenance costs at 3.0 percent for all cases, corresponding to the average of actual data based on data submitted by operators.

Other assumptions presented in Table 6.2 generally agree with previous studies, except that cost of chemicals, utilities, and labor have been indexed to current levels. (See footnotes, Table 6.2.)

6.3.2 Results of Cost Comparison

Table 6.3 presents the line item cost estimates for the models discussed in Section 6.1 for interest rates of 10, 15, and 20 percent. Table 6.4 compares the costs, pollutant removal rates, and cost-effectiveness of control as expressed in dollars per ton of sulfur dioxide removed. All discussion herein will assume a 10 percent interest rate.

Table 6.3 demonstrates the economics of scale of sulfur plant operations. Generally, the most important cost, that of the cost of capital, increases fractionally with increased size.

Maintenance and repair, plant overhead, and other nonlabor operational costs show similar economics of scale, while direct labor costs are practically fixed regardless of plant size. Labor is, however, related to the number of unit operations controlled; therefore, addition of a tail gas treater effectively doubles the labor requirement.

Credits for steam, condensate, and sulfur play a large role in determining the economic viability of a sulfur plant. Since these credits are a direct function of plant size (for a given H_2S/CO_2 acid gas feed), the profit margin is heavily favored for increasing plant size.

Table 6.4 illustrates that a 10.16 Mg/D plant operates at a deficit even without tail gas controls. Tail gas controls turn a highly profitable 50.8 Mg/D plant into a break-even venture, while at 101.6 Mg/D, the tail gas treater halves the profits, but the system still returns a substantial annual surplus.

Cost-effectiveness of tail gas control indicates a similar trend, showing typically \$2,125 per Mg SO_2 cost at 10.16 Mg/D, \$880/Mg at 50.8 Mg/D, and \$675/Mg at 101.6 Mg/D. Interpolating these figures to the current NSPS cutoff at 20.32 Mg/D indicates that the maximum cost per megagram currently incurred (in 1982 dollars) is about \$1,430/Mg, while the more typical cost of a new facility greater than 100 LT/D is considerably less than \$900/Mg. (See Figure 6-1.)

Table 6.3. LINE ITEM COSTS FOR MODEL PLANTS

MODEL 1A (10.16 Mg/d)

Capital cost - $2.54 x 10^6

Direct operating cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 6,395	$ 6,395	$ 6,395
2. treated boiler feedwater	21,615	21,615	21,615
3. electric power	21,210	21,210	21,210
4. fuel gas	17,640	17,640	17,640
5. catalyst	655	655	655
B. Labor			
1. Operators	$84,680	$84,680	$84,680
2. Supervision	41,170	41,170	41,170
C. Maintenance and Repair	$76,200	$76,200	$76,200
D. Supplies and laboratory charges	$16,940	$16,940	$16,940
Fixed Charges:			
A. Capital	$434,490	$333,960	$543,105
B. Taxes	25,400	25,400	25,400
C. Insurance	15,240	15,240	15,240
Plant Overhead:	$40,220	$40,220	$40,220
General Expenses			
A. Administrative	$ 8,020	7,160	9,100
B. Distribution and sales	$ 8,020	7,160	9,100
Total Annualized Costs	$817,895	715,645	928,670
Credits			
1. 1,960 Kp steam	$ 87,320	$ 87,320	$ 87,320
2. 106 Kp steam	5,670	5,670	5,670
3. steam condensate	8,558	8,558	8,558
4. sulfur	399,420	399,420	399,420
Total Credits	$499,265	$499,265	$499,265
Net Annual Operating Cost for Case 1A	$320,439	$218,189	$431,214

Table 6-3. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 1B (10.16 Mg/d)

Capital Cost - $\$4.96 \times 10^6$

Direct operating cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 7,125	$ 7,125	$ 7,125
2. 352 Kp steam	87,615	87,615	87,615
3. treated boiler feedwater	36,040	36,040	36,040
4. electric power	38,010	38,010	38,010
5. fuel gas/hydrogen	39,395	39,395	39,395
6. cooling water	3,190	3,190	3,190
7. catalyst	3,050	3,050	3,050
8. chemicals	810	810	810
B. Labor			
1. Operators	$169,360	169,360	169,360
2. Supervision	82,340	82,340	82,340
C. Maintenance & Repair	$148,800	$148,800	$148,800
D. Supplies & Lab Charges	$ 33,870	$ 33,870	$ 33,870
Fixed Charges			
A. Capital	$848,460	652,140	1,060,545
B. Taxes	49,600	49,600	49,600
C. Insurance	29,760	29,760	29,760
Plant Overhead:	$ 79,540	79,540	79,540
General Expenses			
A. Administrative	$ 16,550	14,650	18,750
B. Distribution and sales	16,550	14,650	18,750
Total Annualized Costs	$1,690,065	1,489,945	1,906,550
Credits			
1. 1,960 Kp steam	$ 90,890	90,890	90,890
2. 106 Kp steam	6,050	6,050	6,050
3. steam condensate	30,945	30,945	30,945
4. sulfur	419,580	419,580	419,580
Total Credits	$547,465	$547,465	$547,465
Net Annual Operating Cost for Case 1B	$1,142,600	$942,480	$1,359,085

Table 6.3. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 2A (50.8 Mg/D)

Capital Cost - $\$4.33 \times 10^6$

	i = 15%	i = 10%	i = 20%
Direct Operating Cost			
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 53,290	$ 53,290	$ 53,290
2. treated boiler feedwater	155,310	155,310	155,310
3. electric power	52,500	52,500	52,500
4. fuel gas	88,200	88,200	88,200
5. catalyst	4,005	4,005	4,005
B. Labor			
1. Operators	84,680	84,680	84,680
2. Supervision	41,170	41,170	41,170
C. Maintenance and Repair	129,900	129,900	129,900
D. Supplies and Lab Charges	16,940	16,940	16,940
Fixed Charges			
A. Capital	740,690	569,310	925,840
B. Taxes	43,300	43,300	43,300
C. Insurance	25,980	25,980	25,980
Plant Overhead	53,645	53,645	53,645
General Expenses			
A. Administrative	15,000	13,300	16,850
B. Distribution and Sales	10,000	13,300	16,850
Total Annualized Costs	$1,519,610	1,344,830	1,708,460
Credits			
1. 1,960 Kp steam	$425,250	$425,250	$425,250
2. 352 Kp steam	15,940	15,940	15,940
3. 106 Kp steam	23,435	23,435	23,435
4. steam condensate	46,200	46,200	46,200
5. sulfur	2,028,600	2,028,600	2,028,600
Total Credits	$2,539,425	2,539,425	2,539,425
Net Annual Operating Cost for Case 2A	($1,019,815)	(1,194,595)	(830,965)

Table 6.3. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 2B (50.8 Mg/D)

Capital Cost - $7.83 x 10^6

Direct Operating Cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 55,965	$ 55,965	$ 55,965
2. 352 Kp steam	284,485	284,485	284,485
3. treated boiler feedwater	172,770	172,770	172,770
4. electric power	95,340	95,340	95,340
5. fuel gas/hydrogen	161,000	161,000	161,000
6. cooling water	10,960	10,960	10,960
7. catalyst	16,290	16,290	16,290
8. chemicals	3,990	3,990	3,990
B. Labor			
1. Operators	169,360	169,360	169,360
2. Supervision	82,340	82,340	82,340
C. Maintenance & Repair	234,900	234,900	234,900
D. Supplies & Lab Charges	33,870	33,870	33,870
Fixed Charges			
A. Capital	1,339,400	1,029,490	1,674,210
B. Taxes	78,300	78,300	78,300
C. Insurance	46,980	46,980	46,980
Plant Overhead	101,065	101,065	101,065
General Expenses			
A. Administrative	27,750	24,650	31,100
B. Distribution & Sales	27,750	24,650	31,100
Total Annualized Cost	2,842,150	2,526,040	3,183,660
Credits			
1. 1,960 Kp steam	439,310	439,310	439,310
2. 106 Kp steam	24,005	24,005	24,005
3. steam condensate	123,345	123,345	123,345
4. sulfur	2,097,900	2,097,900	2,097,900
Total Credits	$2,684,560	$2,684,560	$2,684,560
Net Annual Operating Cost for Case 2B	$ 157,590	($158,520)	$ 499,100

Table 6.3. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 3A (101.6 Mg/D)

Capital cost - $\$6.26 \times 10^6$

Direct Operating Cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. treated boiler feedwater	$402,575	$402,575	$402,575
2. electric power	89,040	89,040	89,040
3. fuel gas	176,400	176,400	176,400
4. catalyst	8,010	8,010	8,010
B. Labor			
1. Operators	84,680	84,680	84,680
2. Supervision	41,170	41,170	41,170
C. Maintenance & Repair	187,800	187,800	187,800
D. Supplies & Lab Charges	16,940	16,940	16,940
Fixed Charges			
A. Capital	1,070,835	823,065	1,338,515
B. Taxes	62,600	62,600	62,600
C. Insurance	37,560	37,560	37,560
Plant Overhead	68,120	68,120	68,120
General Expenses			
A. Administrative	22,460	19,980	25,135
B. Distribution & Sales	22,460	19,980	25,135
Total Annualized Costs	$2,290,650	2,037,890	2,563,680
Credits			
1. 4,300 Kp steam	280,140	280,140	280,140
2. 1,960 Kp steam	92,460	92,460	92,460
3. 352 Kp steam	291,730	291,730	291,730
4. 106 Kp steam	46,870	46,870	46,870
5. steam condensate	35,910	35,910	35,910
6. sulfur	4,057,200	4,057,200	4,057,200
Total Credits	$5,604,310	5,604,310	5,604,310
Net Annual Operating Cost for Case 3A	($3,313,660)	($3,566,420)	($3,040,630)

Table 6.3. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 3B (101.6 Mg/D)

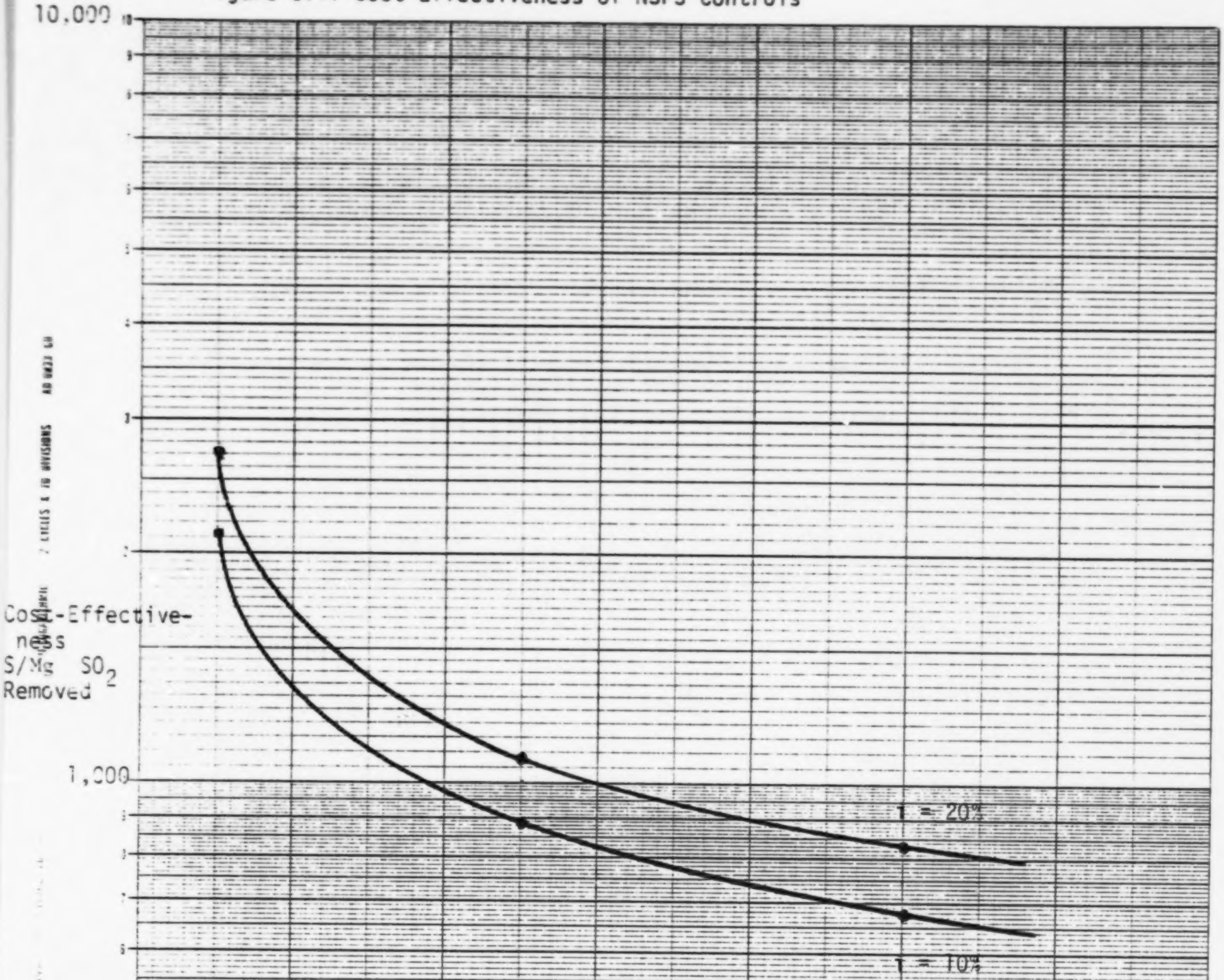
Capital cost - $10.60 x 10^6

Direct Operating Cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. 352 Kp steam	$301,005	$301,005	$301,005
2. treated boiler feedwater	414,820	414,820	414,820
3. electric power	159,600	159,600	159,600
4. fuel gas/hydrogen	363,090	363,090	363,090
5. cooling water	21,925	21,925	21,925
6. catalyst	32,580	32,580	32,580
7. chemicals	7,980	7,980	7,980
B. Labor			
1. Operators	169,360	169,360	169,360
2. Supervision	82,340	82,340	82,340
C. Maintenance & Repair	318,000	318,000	318,000
D. Supplies & Lab Charges	33,870	33,870	33,870
Fixed Charges			
A. Capital	1,813,235	1,393,690	2,266,490
B. Taxes	106,000	106,000	106,000
C. Insurance	63,600	63,600	63,600
Plant Overhead	121,840	121,840	121,840
General Expenses			
A. Administrative	39,750	35,550	44,280
B. Distribution & Sales	39,750	35,550	44,280
Total Annualized Cost	$4,088,745	3,660,800	4,551,060
Credits			
1. 4,300 Kp steam	289,275	289,275	289,275
2. 1,960 Kp steam	921,940	921,940	921,940
3. 106 Kp steam	48,385	48,385	48,385
4. steam condensate	187,215	187,215	187,215
5. sulfur	4,195,800	4,195,800	4,195,800
Total Credits	$5,642,615	5,642,615	5,642,615
Net Annual Operating Cost for Case 3B	($1,553,870)	($1,981,815)	($1,091,555)

Table 6.4. COST & COST-EFFECTIVENESS OF NSPS CONTROLS

	Plant Size, LT/D		
i = 10 percent	10	50	100
Base Case Annual Cost, $	218,189	($1,194,595)	($3,566,420)
Base Case SO_2 Removed, tons/yr	6,765.74	34,362.3	68,724.5
NSPS Case Annual Cost, $	$942,480	($158,520)	($1,981,815)
NSPS Case SO_2 Removed, tons/yr	7,107.22	35,536.1	71,072.2
Cost-Effectiveness, $/ton	$2,126	$882	$674
i = 15 percent			
Base Case Annual Cost, $	320,439	($1,019,815)	($3,313,660)
Base Case SO_2 Removed, tons/yr	6,765.74	34,362.3	68,724.5
NSPS Case Annual Cost, $	$1,142,600	$157,590	($1,553,870)
NSPS Case SO_2 Removed, tons/yr	7,107.22	35,536.1	71,022.2
Cost-Effectiveness, $/ton	$2,413	$1,002	$749
i = 20 percent			
Base Case Annual Cost, $	$431,214	($830,965)	($3,040,630)
Base Case SO_2 Removed, tons/yr	6,765.24	34,362.3	68,724.5
NSPS Case Annual cost, $	$1,359,085	$499,100	($1,091,555)
NSPS Case SO_2 Removed, tons/yr	7,109.22	35,536.1	71,072.2
Cost-Effectiveness, $/ton	$2,723	$1,133	$829

Figure 6.1. Cost-Effectiveness of NSPS Controls



5.4. REFERENCES

1. "SO_2 Emissions in Natural Gas Production Industry--Background Information for Proposed Standards," EPA 450/3-82-023a, January 1983, Chapters 6 and 8.

2. Sulfur Recovery Study - Onshore Sour Gas Production Facilities, Ralph M. Parsons Company, August 20, 1981.

3. Standards Support and Environmental Impact Statement Volume 1: Proposed Standards of Performance for Petroleum Refinery Sulfur Recovery Plants, U.S. EPA, September 1976. Chapter 8.

7. OTHER IMPACTS REVIEWED

7.1 NON-AIR ENVIRONMENTAL IMPACTS

7.1.1 Water Pollution Impact

Of the control technologies examined which can meet NSPS requirements, little if any impact upon water quality is foreseen. The amine absorption/ regeneration processes generate significant quantities of reusable process water normally filtered and sent to the sour water stripper. Only if significant SO_2 breakthrough occurs, does the water form soluble sulfates and sulfites, in which case the water may be sent to the plant water treatment facility. For an integrated refinery, this would represent substantially less than 1 percent of total water treated. It is presumed that this condition occurs infrequently, based on results of EPA's survey.[1,2,3]

The oxidation process does produce process water containing dissolved sulfates; however, this process is not planned on any NSPS units at this time.[4]

The reduction/Stretford process should produce identical sour water streams as the amine absorption process. The vendor of this process recommends two-stage quench towers, ensuring that only small amounts of water require treatment for sulfites/sulfates, with the majority reporting to the sour water strippers for re-use.[5]

The Stretford process itself can become a potential source of water pollution, since by-product sulfates and thiosulfates require periodic purging. Disposal methods of this purge stream involve recovery of sodium value by evaporation or spray drying, biological degradation, or oxidative combustion.[6] After salt recovery, the solid residue may be landfilled. The next section discusses another alternative which results in no liquid waste purge.

7.1.2 Solid Waste Impacts

The potential solid wastes from NSPS control systems consist of spent reduction catalysts (cobalt-molybdenum) and solid residue from Stretford purge systems. The spent catalysts have market value and have historically been returned to the vendors for credit when replaced. One recent study concludes that spent Stretford solution residues are very small in volume and have an insignificant solid waste impact.[7] Another opinion, however, is that any solid waste, no matter how small, presents disposal problems in some locations. The vendor for this system indicates that an alternative sulfur recovery step is now available which will not require purge and disposal of the absorbing solution.[8]

The conclusion is that NSPS controls may precipitate a minor solid waste problem, but in the near future may diminish as new operations choose waste-free technologies.

7.2 ENERGY AND ENERGY-RELATED IMPACTS

The most significant negative impact of applying tail gas treatment results from the additional steam, hydrogen, electricity, and fuel gas consumed. In all processes examined capable of achieving NSPS levels, low pressure (352 kilopascal) steam and electricity are consumed. Fuel gas consumption is also significant where final incineration is required; however, the reduction/Stretford option results in fuel gas savings. Hydrogen consumption depends upon Claus operation and Claus feed characteristics; in some cases, no hydrogen is consumed while others require nominal amounts of hydrogen.[9,10,11]

For the 101.6 Mg/yr model plant, incremental annual energy consumption (NSPS case less the Base Case) is as follows:

electricity	1.411×10^{6} KWH or 5.08×10^{12} joule (j)
fuel gas/hydrogen	56.22×10^{12} j
9,300 Kp steam	$(1.60 \times 10^{12}$ j)
1,760 Kp steam	$(5.53 \times 10^{12}$ j)
352 Kp steam	127.56×10^{12} j)
106 Kp steam	$(1.12 \times 10^{12}$ j)
Net Consumption:	180.61×10^{12} j/yr

Since the sulfur plant emission controls account for an annual reduction of 2,316.2 Mg/y (2,552.45 t/y), the energy cost is about 78×10^9 joule per Mg SO_2 removed. The secondary impact of energy consumption, air emissions generated to replace energy loss, may be calculated based on a coal-fired utility boiler assumption. This worst-case scenario indicates that the 78×10^9 joule of coal heating value expended to convert one megagram of SO_2 into one-half megagram of salable sulfur would generate .045 Mg SO_2, .001 Mg particulate matter, .002 Mg NO_x, and 0.25 Mg of solid waste.

7.3 OTHER IMPACTS

The only other impacts of significance incurred by NSPS controls involve the additional labor requirements and the overall reliability of sulfur plant operations. In Chapter 6, a 2/3 man-per-shift incremental impact was assigned for addition of tail gas controls. In actuality, the sulfur recovery unit would likely already have two operators per shift assigned to the amine and Claus units. Addition of a tail gas unit would be integrated into the control scheme such that the two operators would devote one-third of their time to tail gas controls and, therefore, less time to their other responsibilities. This would likely require more reliance on automated controls for other processes and improved data retrieval and storage at the control panel. These phenomena are in fact taking place as sulfur recovery areas undergo replacement and expansions of existing facilities.[12]

Reliability of the sulfur plant is typically 95 percent at new tail gas installations; however, for the older tail gas installations, reports indicate reliabilities of near 100 percent and maintenance costs less than or equal to Claus plant levels.[13,14,15] Hence, for the facilities modelled in this study, it can be argued that reliability overall could not have decreased more than 5 percent. In fact, the Claus/tail gas failures often occur together, thus, the conclusion is that reliability of a properly designed and operating tail gas unit does not significantly impact sulfur plant operations.

Overall, the impact of tail gas controls on refinery operations is a reworking of operator schedules to include 1/3 time per operator devoted to tail gas controls, and a near doubling of anticipated maintenance labor on the sulfur plant, the majority of which would occur simultaneously for Claus and tail gas treaters.

7.4 REFERENCES

1. Confidential letter, E. P. Crockett, American Petroleum Institute, to C. B. Sedman, U.S. EPA, dated June 30, 1982.

2. Sedman, C. B., U.S. EPA, Trip Report - ARCO Refinery, Houston, Texas, dated September 20, 1982.

3. Letter, C. M. Tyler, Standard Oil Company of Ohio, to Don Goodwin, U.S. EPA, dated July 15, 1982.

4. Letter, D. H. Dilworth, Davy-McKee, to C. B. Sedman, U.S. EPA, dated October 5, 1982.

5. Telephone conversation, C. B. Sedman, EPA, and R. E. Warner, R. M. Parsons Company, October 19, 1982.

6. "SO_2 Emissions in Natural Gas Production Industry - Background Information For Proposed Standards", EPA 450/3-82-023a, January 1983, pp. 7-9 to 7-12.

7. Reference 6.

8. Letter, J. C. Brocoff, R. M. Parsons Co., to S. T. Cuffe, U.S. EPA, February 16, 1983.

9. Reference 2.

10. Sedman, C. B., U.S. EPA, Trip Report - Phillips Petroleum Refinery - Sweeny, Texas, dated September 27, 1982.

11. Sedman, C. B., U.S. EPA, Trip Report - Mobil Oil Refinery - Beaumont, Texas, dated October 15, 1982.

12. Reference 10.

13. Reference 1.

14. Reference 3.

15. Confidential letter, G. E. Lowe, Marathon Petroleum Company, to D. R. Goodwin, U.S. EPA, dated September 16, 1982.

8. RECOMMENDATIONS

8.1 REVISIONS TO NSPS

8.1.1 Sulfur Emissions

From the previous chapters, it is shown that the only significant disadvantage of requiring NSPS controls is cost, both capital and operating. Capital costs are essentially doubled to remove the final four percent of potential SO_2 emissions. Operating costs are essentially doubled, since labor, maintenance, and cost of capital are doubled. Steam and sulfur credits are not significantly affected.

Potential revisions to the standard could include lowering allowable emissions to, say 125 ppmv, or relaxing the requirements to 500, 1,000, or 1,500 ppmv (corrected to zero percent oxygen). Raising or lowering to the above levels would accomplish very little from a cost standpoint, since the same systems as found in NSPS application would be used.[1] Therefore, capital expenditures would not be significantly affected and only the energy portion (and possibly maintenance costs) of operating costs would be noticeably affected.[2,3]

To make a significant impact on capital and operating cost, the NSPS would either have to be revised to allow the Claus extension processes, or dropped altogether. Claus extension processes have not been subjected to modelling and analysis, but current experience indicates that the typical control level is 98.6 percent efficiency.[4] Hence, for a 101.6 Mg/d facility, the additional operating cost would be about $578,000 for a cost-effectiveness of $395/Mg SO_2 removed. The Claus plant would remain a major SO_2 source, emitting nearly 1,000 megagrams SO_2 per year. With full tail gas control at $750/Mg, the facility emits less than 100 megagrams SO_2 annually and could be considered less than a major emission source.

A problem not mentioned in this study surfaced during the review of this document in draft form. Briefly, the NSPS assumes all sulfur species in incinerators to be converted to SO_2; hence, only SO_2 is regulated.

One State agency has commented that temperature and O_2 monitoring of incinerators are needed to ensure total sulfur oxidation to SO_2. It is recommended that the EPA pursue this problem in conjunction with other potential changes to be discussed.

8.1.2 Lower Capacity Cut-off

Another way of reducing costs of NSPS would be to raise the lower capacity exemption of 20.12 Mg/D to some other level, say 50.8 Mg/D. As shown back in Chapter 4, Table 4.2., only 3 of 24 planned units are in the 20 to 50 Mg/D range. Additionally, Chapter 6, Figure 6.1 suggests that the cost-effectiveness at 20.32 Mg/D is not significantly different at 50 Mg/D. Only at less than 10 LT/D capacities do the cost-effectiveness curves become steep enough to convincingly serve as an economic basis for less stringent regulations. Unless some arbitrary cost-effectiveness value is chosen as a guide for determining regulatory levels, the recommended path is to retain the 20.32 Mg/D capacity cut-off.

8.1.3 Other Emissions

Since most sulfur plants are subject to State and local regulations, emission tests are frequently conducted for other pollutants such as carbon monoxide, particulate matter, nitrogen oxides, and hydrocarbons. No specific control techniques for these pollutants have been identified, so it is assumed that the basis for regulation is good operation of the process. Examination of emission test results shows that emission levels of nonsulfur species other than carbon monoxide are well below the NSPS sulfur level. Table 8.2 contains these emissions and suggests that regulation of other sulfur plant emissions are not warranted on a national basis.

Table 8.2. TYPICAL SULFUR PLANT EMISSIONS WITH TAIL GAS CONTROL[5,6]

	With Incineration	Without Incineration
CO ppmv	650	300
CH_4 ppmv	--	55
SO_2 ppmv	86	<1
H_2S ppmv	--	9
particulate gr/DSCF	<.0002	--

8.2 REVISIONS TO MONITORING REQUIREMENTS

8.2.1 Total Sulfur Monitors

Although monitoring specifications have not been made for monitors under the sulfur plant NSPS, several total sulfur monitors have been reported on refinery sulfur plants.[7,8] To date, performance of these monitors has been less than satisfactory to the operators, although many problems pertain to sample collection and conditioning.[8] Since sample collection problems can normally be solved, a further investigation of total sulfur monitors seems warranted with the goal of developing performance specifications to complement the monitoring requirements of the NSPS.

8.2.2 Hydrogen Sulfide Monitors

Monitors specifically for hydrogen sulfide are essentially the same type (lead acetate tape - light dispersion) as observed during preparation of the NSPS.[9,10] Since the state-of-the-art for H_2S monitors has apparently not advanced since the NSPS, it would seem expedient to investigate H_2S monitoring in combination with total sulfur monitoring with the goal of simultaneous monitoring of reduced sulfur and H_2S, just as both are currently measured by EPA Method 15.

8.2.3 Sulfur Dioxide and Oxygen Monitors

Sulfur dioxide monitors are found on many new NSPS facilities where a final incinerator is used for H_2S destruction.[11,12] Most surveyed use an in-stack SO_2 and oxygen monitors similar to that employed on coal-fired utility boilers. The standard currently does not address the need for oxygen monitors to convert SO_2 to an oxygen-free basis. It would appear that specifications can be applied to refinery sulfur plants. It is therefore recommended to amend the sulfur plants NSPS to include oxygen monitoring.

8.3 REVISIONS TO COMPLIANCE TESTING REQUIREMENTS

At some sites, minor modifications to EPA Method 15 have been instituted to alleviate problems in sample collection such as moisture and sulfur accumulation.[13] These problems are generally recognized and approval of modifications by the enforcement authority has been granted.[14]

Method 6 for sulfur dioxide is considered a universally accepted reference method and no change is indicated herein.

8.4 SUMMARY OF RECOMMENDATIONS

Based on costs, cost-effectiveness, and other environmental impacts, the current NSPS including the 20.32 Mg/D lower capacity exemption should be retained. Oxygen monitoring requirements should be added to the NSPS, and total sulfur monitors should be examined to see if specifications based on a reliable system may be developed. Temperature monitoring for incinerators should also be considered to ensure minimal non-SO_2 emissions where only SO_2 emissions are regulated. No other changes to the NSPS appear warranted, save a possible revision to EPA Test Method 15.

8.5 REFERENCES

1. Letter, W. T. Knowles, Shell Oil Company, to Charles B. Sedman, U.S. EPA, dated August 24, 1982.

2. Reference 1.

3. Letter, H. J. Grimes, ARCO Petroleum Products Company, to Charles B. Sedman, U.S. EPA, dated October 5, 1982.

4. Letter, C. V. Rice, Amoco Oil Company, to Charles B. Sedman, U.S. EPA, dated October 18, 1982.

5. Letter, R. M. Thompson, Shell Oil Company, to Charles B. Sedman, U.S. EPA, dated October 12, 1982.

6. Letter, L. C. Worley, Exxon Company, USA, to Charles B. Sedman, U.S. EPA, dated October 14, 1982.

7. Sedman, C. B., U.S. EPA - Trip Report - Phillips Petroleum Refinery, Sweeny, Texas, dated September 27, 1982.

8. Sedman, C. B., U.S. EPA - Trip Report - Mobil Oil Refinery, Beaumont, Texas, dated October 15, 1982.

9. Sedman, C. B., U.S. EPA - Trip Report - Beavon Sulfur Removal Units, dated November 5, 1973.

10. Reference 8.

11. Reference 7.

12. Letter, C. M. Tyler, Standard Oil Company of Ohio, to C. B. Sedman, U.S. EPA, dated July 15, 1982.

13. Confidential letter, R. J. Niederstadt, Mobil Oil Corporation, to Don Goodwin, U.S. EPA, dated June 15, 1982.

14. Telephone conversation, B. Ferguson, Harmon Engineering and Testing, Inc., to C. Sedman, U.S. EPA, dated November 18, 1982.

APPENDIX A
COST ESTIMATING TECHNIQUES AND RESULTS OF COST/ANALYSES FOR SULFUR PLANTS

A.1. CAPITAL COST ESTIMATES

A.1.1 Claus Plants

The most recent work involving capital cost estimates for Claus plants is the 1981 Ralph M. Parsons Company study prepared for The Onshore Gas Production NSPS.[1] Although the study was directed primarily toward lean (<50% H_2S) acid gas streams, the cost estimates allow for reasonable extrapolation to the 80% H_2S refinery case and direct comparison to other data sources. Additional cost estimates were obtained from responses to EPA inquiries via 114 letters and phone calls to facilities having Claus plants subject to the NSPS. Though not directly used, previous cost estimates from the original EPA study on refinery Claus plants (1975) and the GPA Panel discussions in the Oil and Gas Journal were consulted for comparison.[2,3] Since all previous cost studies were performed in English units, English units are used in these appendices for consistency, then converted to metric units in the main report body.

Table A-1 presents the Claus capital cost estimates used to develop model costs. These costs are all indexed to July 1982 dollars using the process industry cost indices from Chemical Engineering:

1974	165.4
June 1975	182.4
1978	218.8
April 1980	257.3
1980	261.2
January 1981	276.6
July 1982	314.2

The Parsons capital estimates in Table A-1 are for 2 or 3 stage Claus plants with thermal oxidizer and stacks selected to give uniform ground level SO_2 concentrations. Some cost estimates for the larger Claus plants also include oxidizer and stack, but with unknown design basis. Figure A-1 is a logarithmic plot of cost data from Table A-1

Table A-1. VARIOUS ESTIMATES OF CLAUS INVESTMENT COSTS

Source of Estimate	Claus Capacity LT/D	Acid Gas H_2S, %	No. of Stages	Estimated Capital Cost $x10^6$ (corrected to July 1982)	Year, Month of Estimate	Comment
Parsons Study	10	50	2	2.50 (2.84)	Jan. 1981	Installed cost, no heat recovery
	10	50	3	2.87 (3.26)	"	"
"	10	20	2	2.95 (3.35)	"	"
"	10	20	3	3.29 (3.74)	"	"
"	10	12.5	2	3.08 (3.50)	"	"
"	10	12.5	3	3.34 (3.79)	"	"
"	100	50	3	6.47 (7.35)	"	Waste heat recovery from thermal oxidizer (incinerator)
"	100	20	3	9.05 (10.28)	"	"
"	100	12.5	3	11.21 (12.73)	"	"
"	555	20	3	26.23 (29.80)	"	"
"	1000	80	3	22.30 (25.33)	"	"
"	1000	50	3	26.10 (29.65)	"	"
EPA Background Document	5	80	3	0.757 (1.30)	June 1975	No heat recovery
	10	80	3	0.902 (1.55)	"	"
	100	80	3	2.783 (4.79)	"	"
OGJ GPA Panel Report	100	80(?)	3(?)	3.5 (5.03)	1978	Assumes typical refinery installation
SOHIO	100	75	3	5.45 (655)	July 1980	With stack heat rec.
Parsons	10.3	80	3	2.07 (2.53)	April 1980	No heat recovery

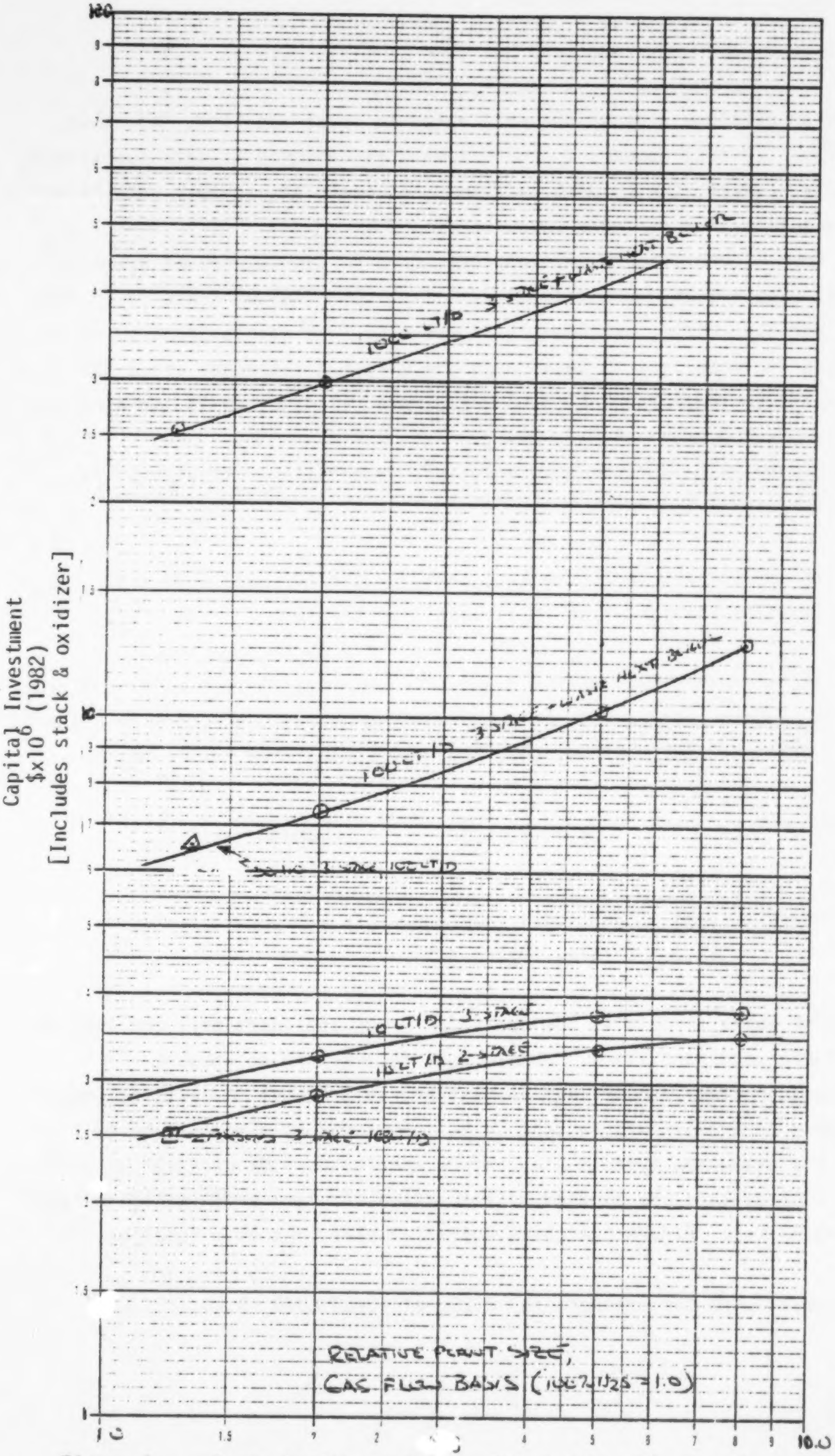


Figure A-1. Claus Capital Costs vs. Plant Size (Gas Flow Basis) based on relative size of the unit based on total gas flow. As shown, the 100 LT/D case appears a good estimate as compared to data supplied by SOHIO[4]; however, the 10 LT/D case does not correlate as well with Parson's own estimate on a refinery case.

From Figure A-1, the extrapolated data for an 80 percent H_2S case (plant size = 1.25) were plotted as a function of Claus capacity in long tons per day (LT/D), as shown in Figure A-2. The capital cost curve to be used for modelling is based upon The Parson's estimates above 100 LT/D and a fit to the Parsons 10.3 LT/D estimate. This curve is for a no heat recovery assumption.

Above 100 LT/D, the cost curves are essentially a straight-line relationship of the form

$$y = mx^{0.6}$$

Referring back to Figure A-1, data for the 100 LT/D case also approximate a straight-line relationship of the form

$$y = mx^{0.4}$$

Therefore, for any sulfur plant of known capital cost (1982 dollars) C_1, of capacity rating LTD_1, and $\%H_2O$ in feed $(H_2S)_1$, the cost of a second Claus plant C_2 with capacity LTD_2 and feed composition $(H_2S)_2$ may be found by:

Equation A-1
$$C_2 = C_1 \left(\frac{LTD_2}{LTD_1}\right)^{0.6*} \left(\frac{(H_2S)_1}{(H_2S)_2}\right)^{0.4}$$

where $100 \leq LTD_1, LTD_2 \leq 1000$
$12.5 \leq (H_2S)_1, (H_2S)_2 \leq 90$

The above formula is obviously for rough estimates only and includes the incinerator and stack. Should heat recovery or an unusual incinerator/stack requirement be desired, adjustments to costs estimated as above or from Figure A-2 should be considered as discussed below.

The estimated typical Claus stack and incinerator capital costs in July 1982 dollars are plotted as a function of plant size (gas flow basis) in Figure A-3. Figure A-4 shows a similar plot, but as a function of

---

* At 10-40 LT/D the exponent is 0.20, at 40-80 LT/D 0.40.

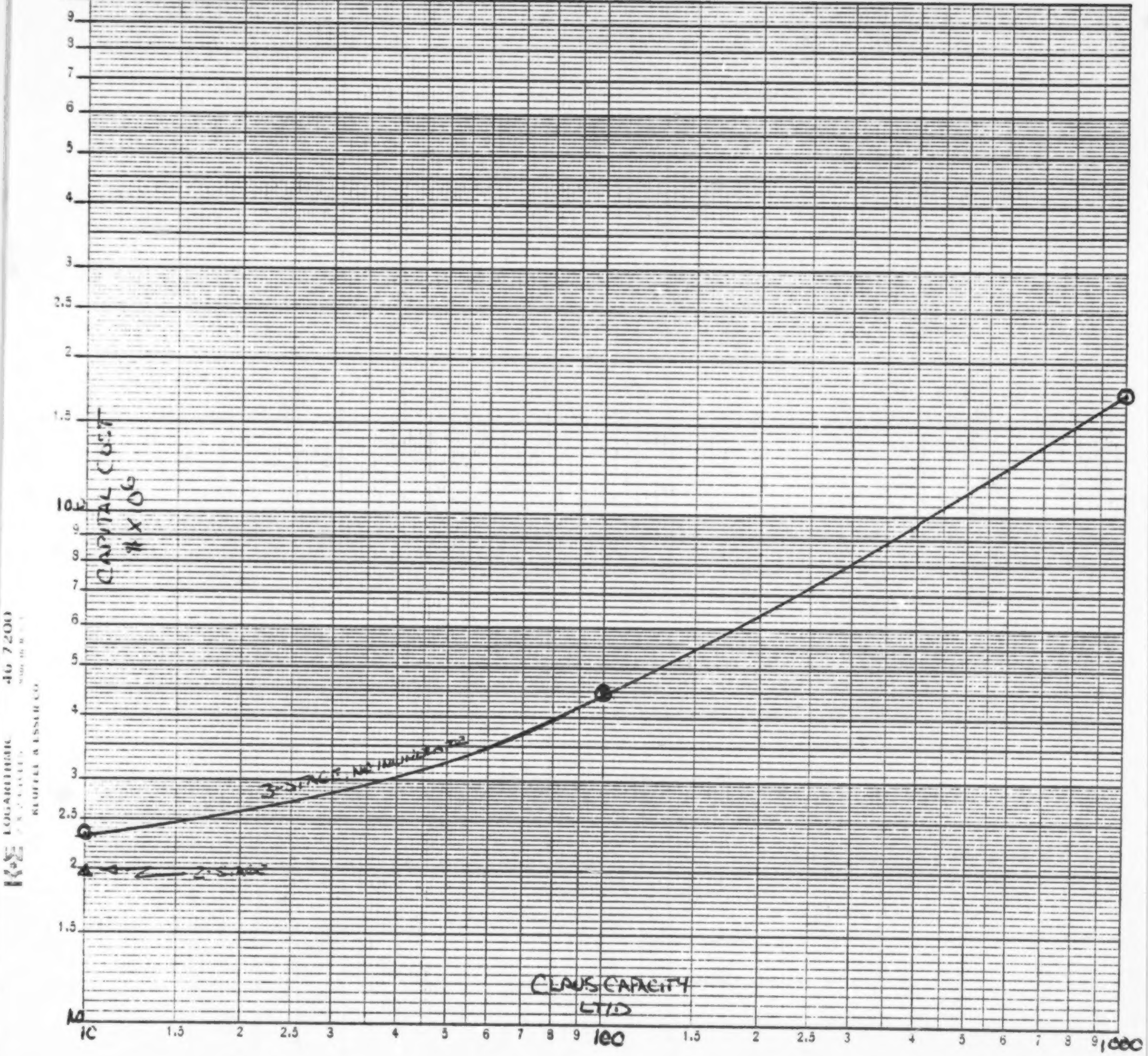


Figure A-2. Claus Only Capital Cost vs.
Plant Sulfur Capacity @ 80% H_2S Feed

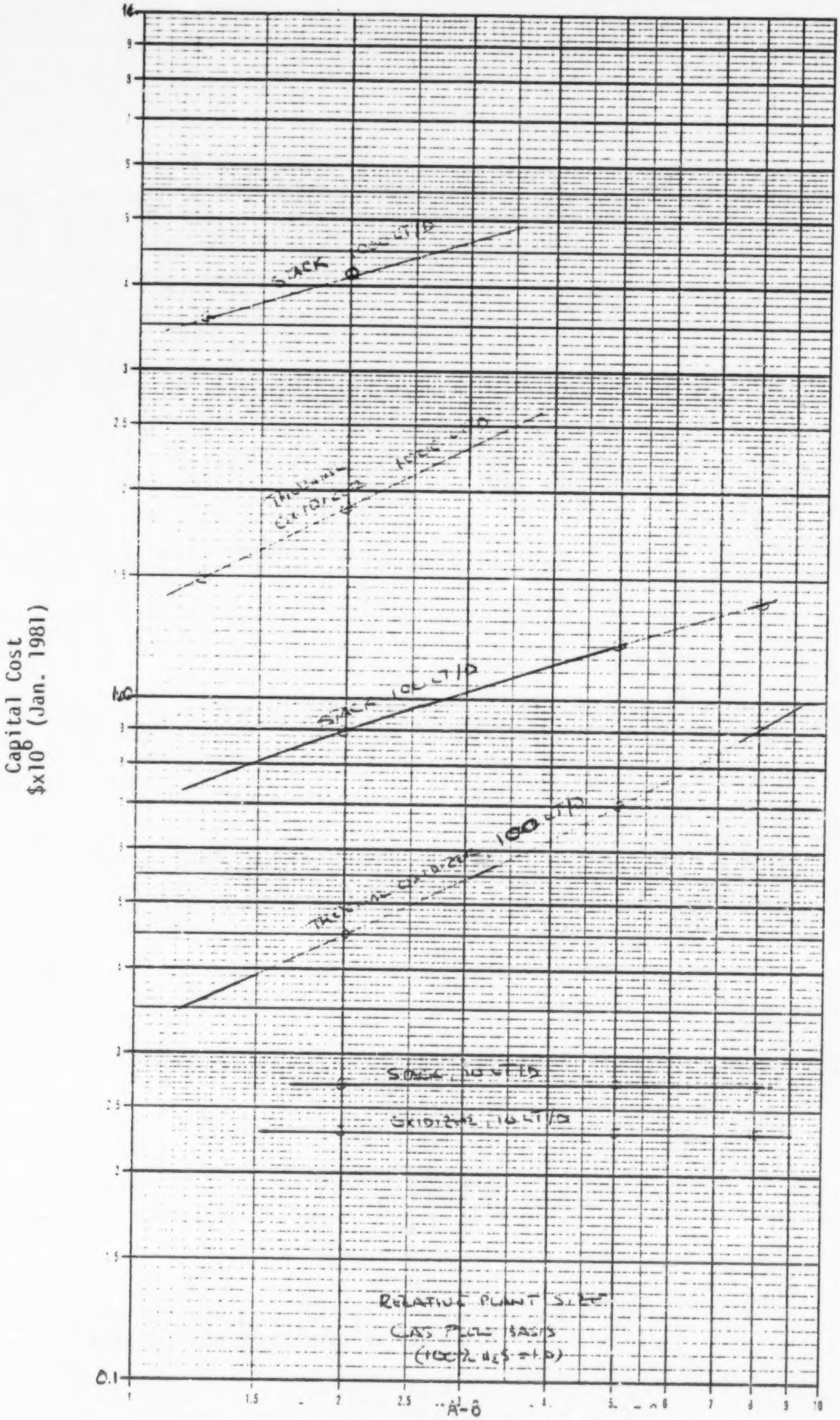


Figure A-3. Stack and Incinerator Capital Costs

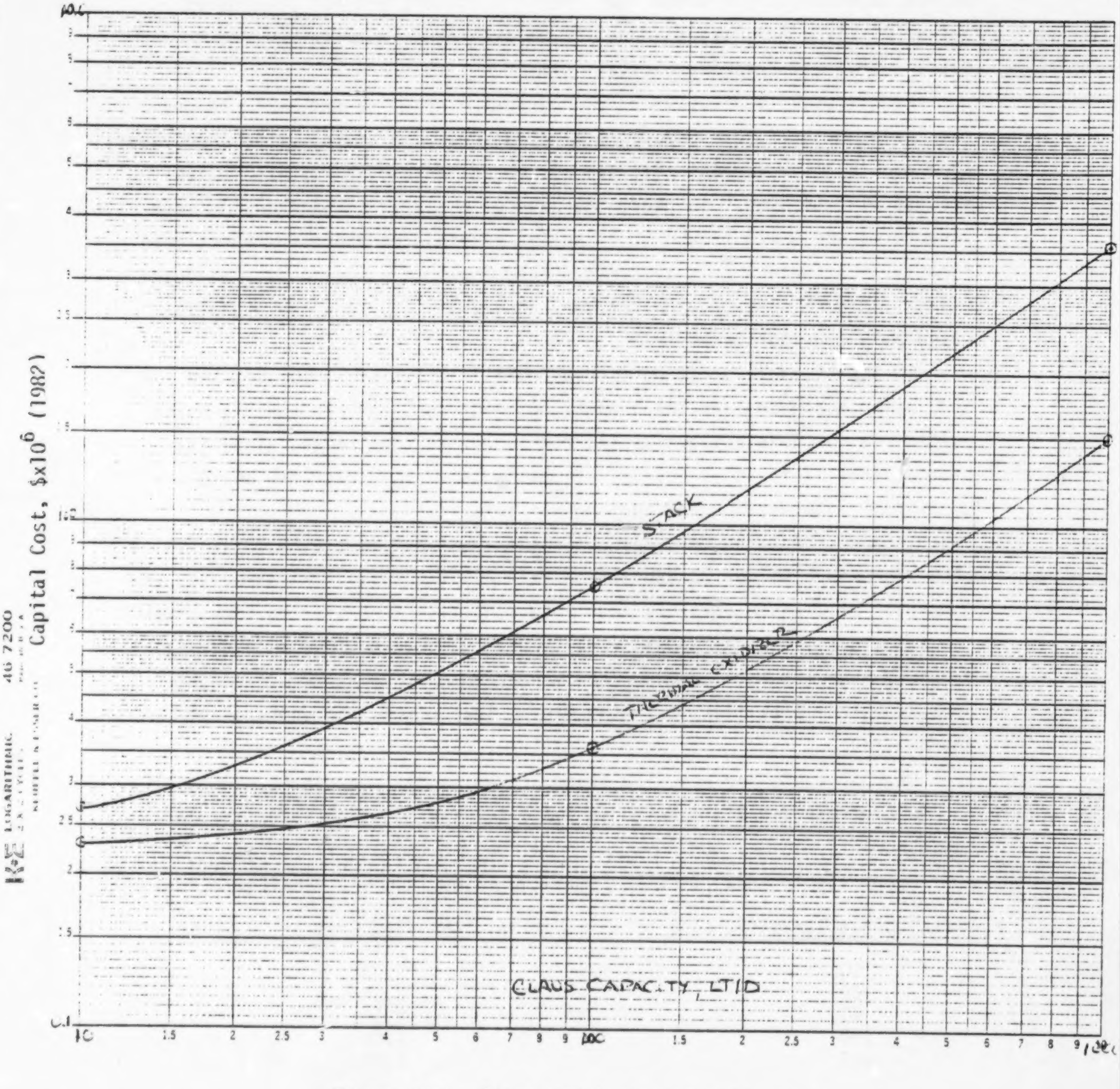


Figure A-4. Stack and Incinerator Costs vs. Claus Capacity (80% H_2S)

sulfur capacity. Figure A-5 plots capital costs of waste heat recovery boilers for plants greater than 100 LT/D. A plot similar to Figure A-4 is not necessary since waste heat recovery boilers are not considered below 100 LT/D, the largest model to be examined.

A.1.2. Tail Gas Treating Capital Costs

As discussed in Chapter 5, the purpose of this report is to assess the impact of NSPS upon Claus plant operation. Therefore, it is unnecessary to evaluate all potential tail gas processes, rather, a representative process will suffice. Further, the area of interest in determining cost impacts is the small (10-50 LT/D) sulfur plant which represent worst-case impacts. Ultimately this analysis should answer the questions, "What are typical control costs?", and "Is the current 20 LT/D capacity exemption reasonable considering costs?"

To answer these questions, three model facilities at 10, 50, and 100 LT/D were chosen to span the area of most interest and provide a 3-point cost curve for possibly evaluating models within this range. Assuming that control costs at 100 LT/D are reasonable, larger facility costs are of minimal interest for the purposes of this study.

Because the amine tail gas process is dominant in the less than 100 LT/D size range (18 of 20 operataing tail gas treaters or 90 percent), it is chosen as a representative model basis. It is important to note that the amine system is not necessarily the lowest cost process in this size range, rather the most common. One vendor of both amine and Stretford processes indicates that the amine may be less costly for units of 30 LT/D and smaller.[5]

Capital costs for actually installed amine tail gas units in the 10 to 100 LT/D range are presented in Table A-2 and adjusted to a July 1982 basis.

Table A-2. CAPITAL COSTS FOR AMINE TAIL GAS TREATERS

Parent Claus Capacity, LT/D	Capital Installed Cost $\$ \times 10^6$ (1982)
10	2.31
20	2.84
60	2.50*
100	4.68
165	5.97

* Thought to be 1978 equipment + 1978-82 construction.

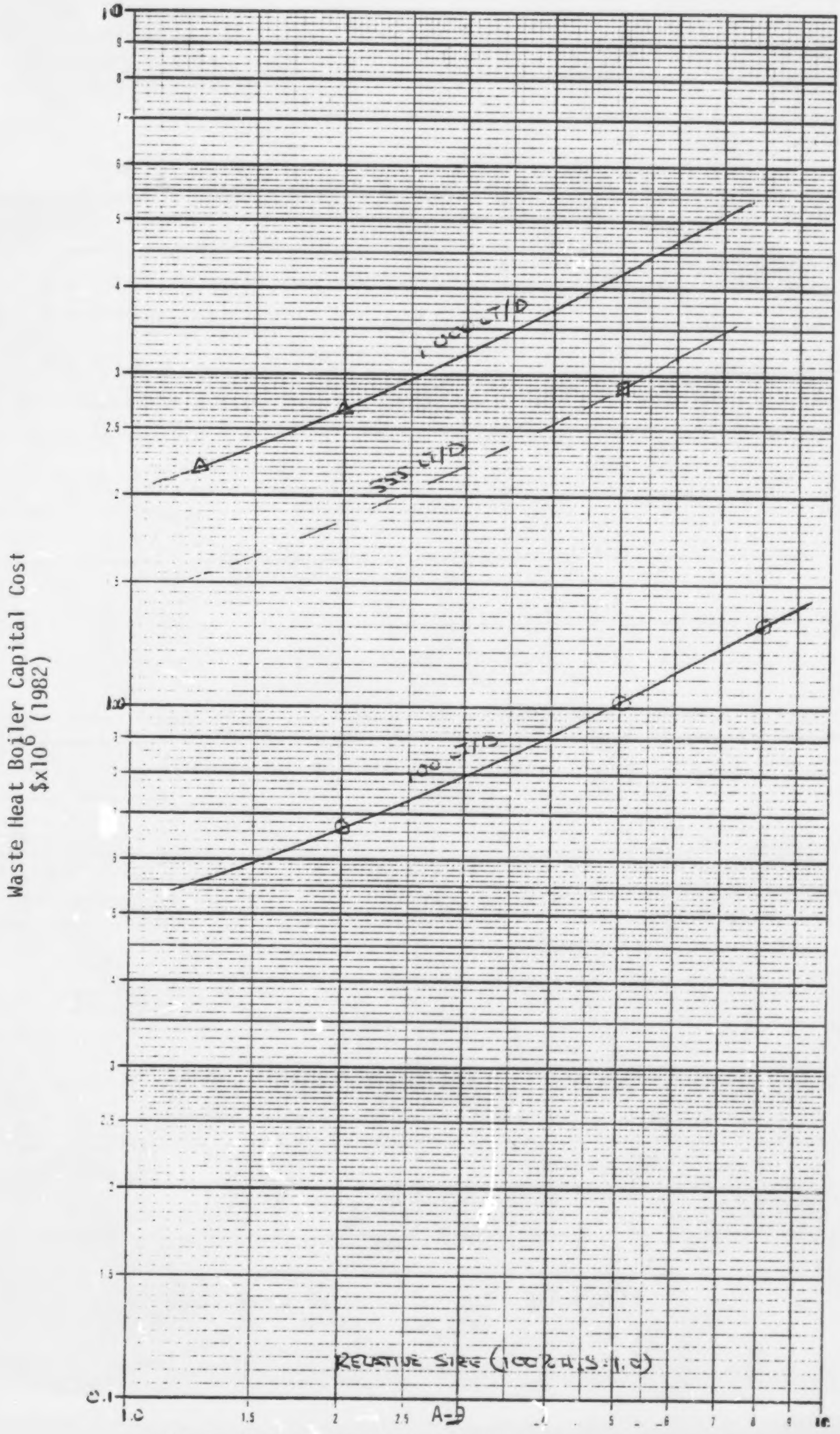


Figure A-5. Waste Heat Recovery System Capital Costs

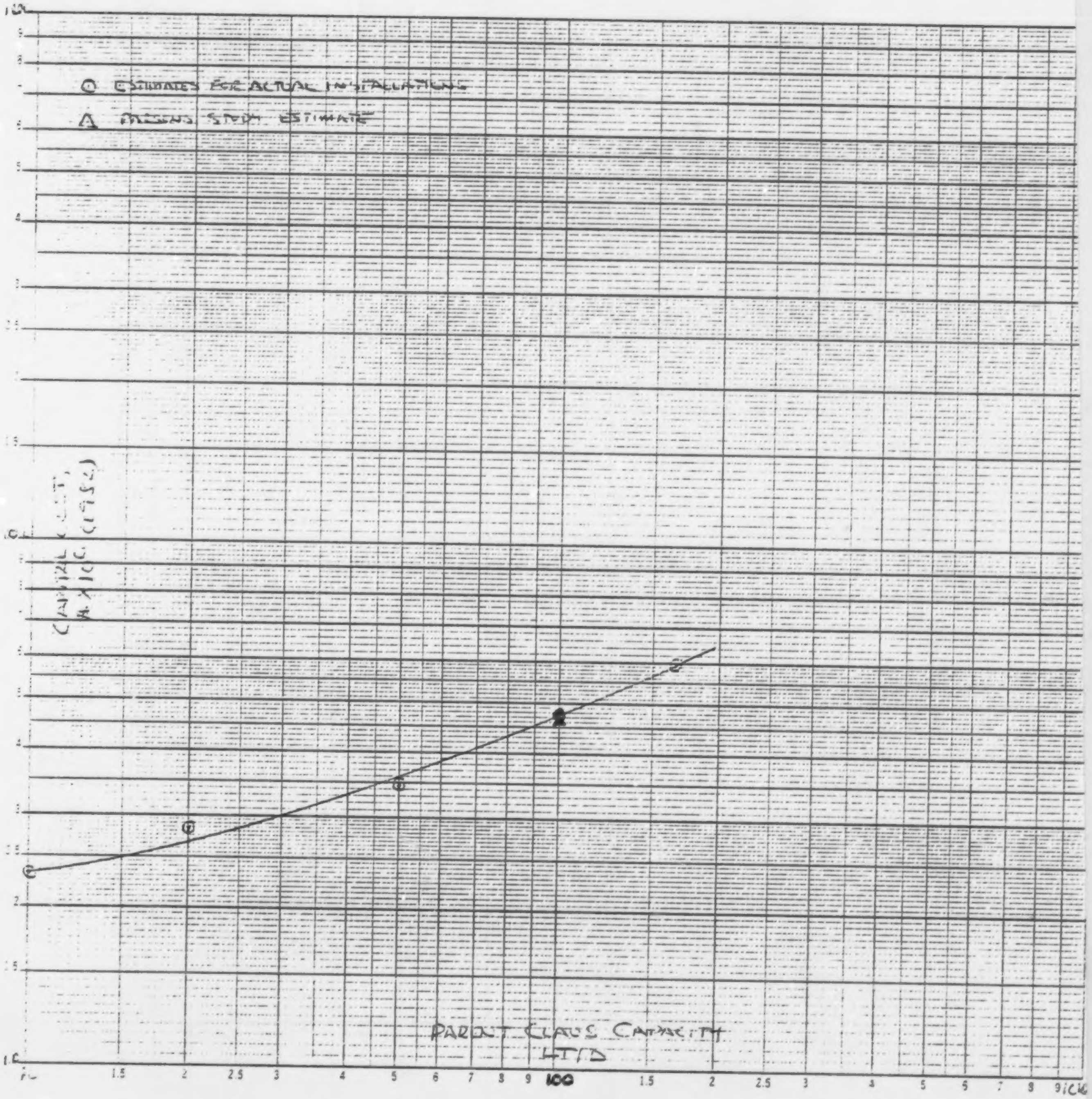


Figure A-6. Capital Costs of Amine-Based
Tail Gas Treaters

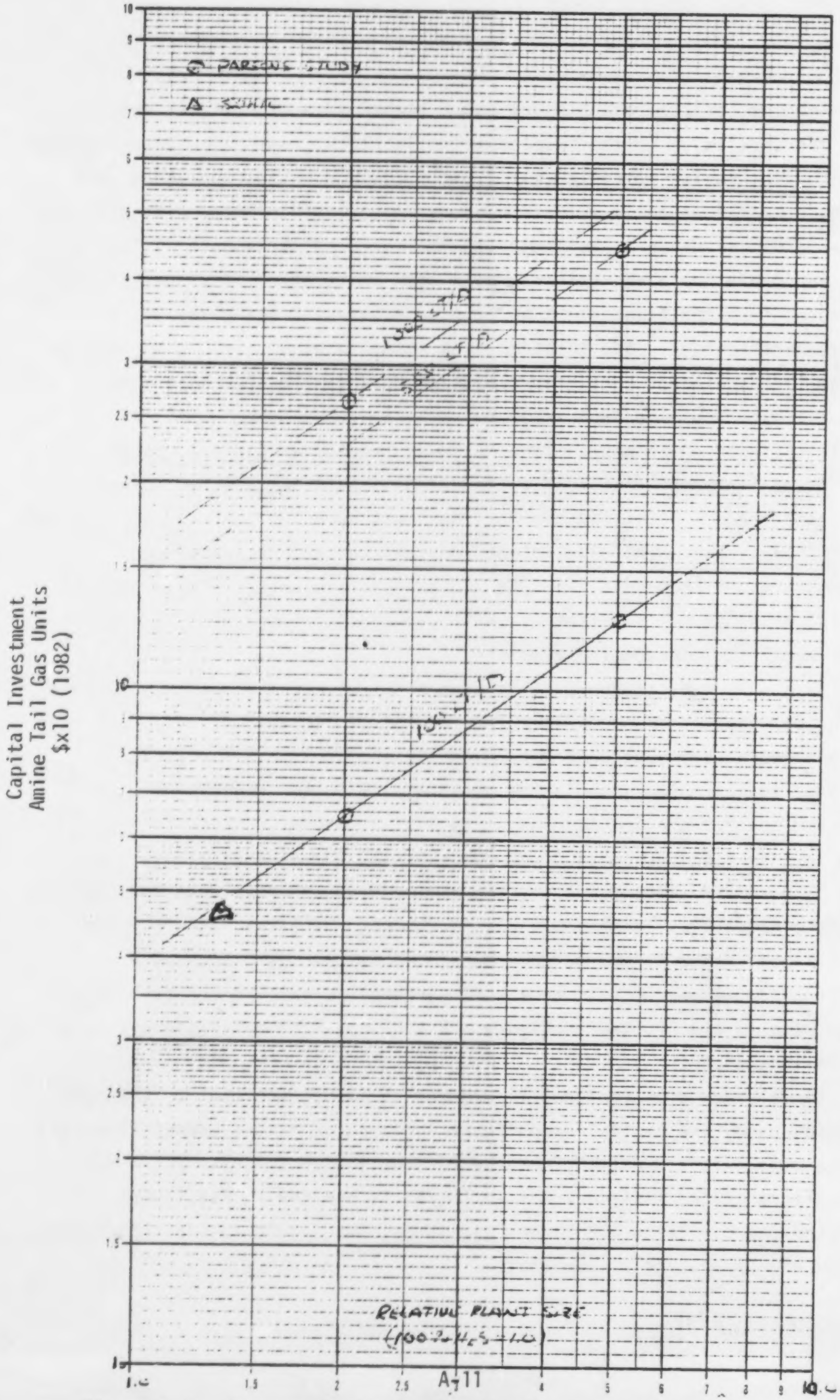


Figure A-7. Parsons Estimates for Capital Costs of Amine Tail Gas Units

The costs in Table A-2 represent a combination of retrofit and new tail gas treaters. In the case of retrofit units, costs have been adjusted down to account for retrofit costs, while for new units, the costs were disaggregated from total sulfur recovery costs.[6] Therefore, a significant degree of uncertainty is reflected in the above costs because no data were available for a new tail gas unit with costs of the tail gas treater separated from the Claus plant and, in some cases, Claus plant amine treater and boilers. The \$2.50 million estimate at 60 LT/D is thought to be the 1978 equipment cost + installation during 1978-82. A reasonable 1982 estimate would be \$3.6 x 10^6.

A.1.3. Effects of Combined Claus/Tail Gas Treater on Capital Costs

To estimate the combined cost of Claus + tail gas treater is not straightforward. First, if the tail gas unit recycles the removed material to the Claus plant, the Claus plant requires increased capacity to accommodate the increased gas flow and sulfur recovery. This increased capital expenditure is offset by the lower capital incurred by a smaller stack required to disperse emissions.

In the Parsons study, the increase in Claus plant expenditure due to amine tail gas testing were \$0.32x10^6 at 100 LT/D, 50% H_2S for a 7.06 percent increase in cost; \$1.08x10^6 at 100 LT/D, 20% H_2S for a 17.3% increase in capital cost. In the model 100 LT/D plant chosen (80% H_2S), the average increase in Claus capacity is 3.3 percent. Also, the gas flow is increased by some 4.27 percent; hence, the percentage H_2S drops from 80 to 78.68 percent. Also, the engineering design allows for doubling of anticipated recycle stream for safe design; therefore, the increased capital cost based on the formula developed earlier is estimated at $[(1.066)^{0.6} (\frac{80}{78.68})^{0.4} - 1]$ or a 4.60 percent increase in capital cost. These results are plotted in Figure A-8 and appear to correlate well with the Parsons study.

Since stack size (height) is assumed to be proportional to the mass emission rate, the capital expenditure for a stack is therefore a function of the mass emission rate. From the Parsons study, the data for stack expenditure versus emission rate in lbs/hr is plotted in Figure A-9 for selected cases. Below 150 lb/hr SO_2, the stack cost is essentially fixed at \$30,670 (July 1982).

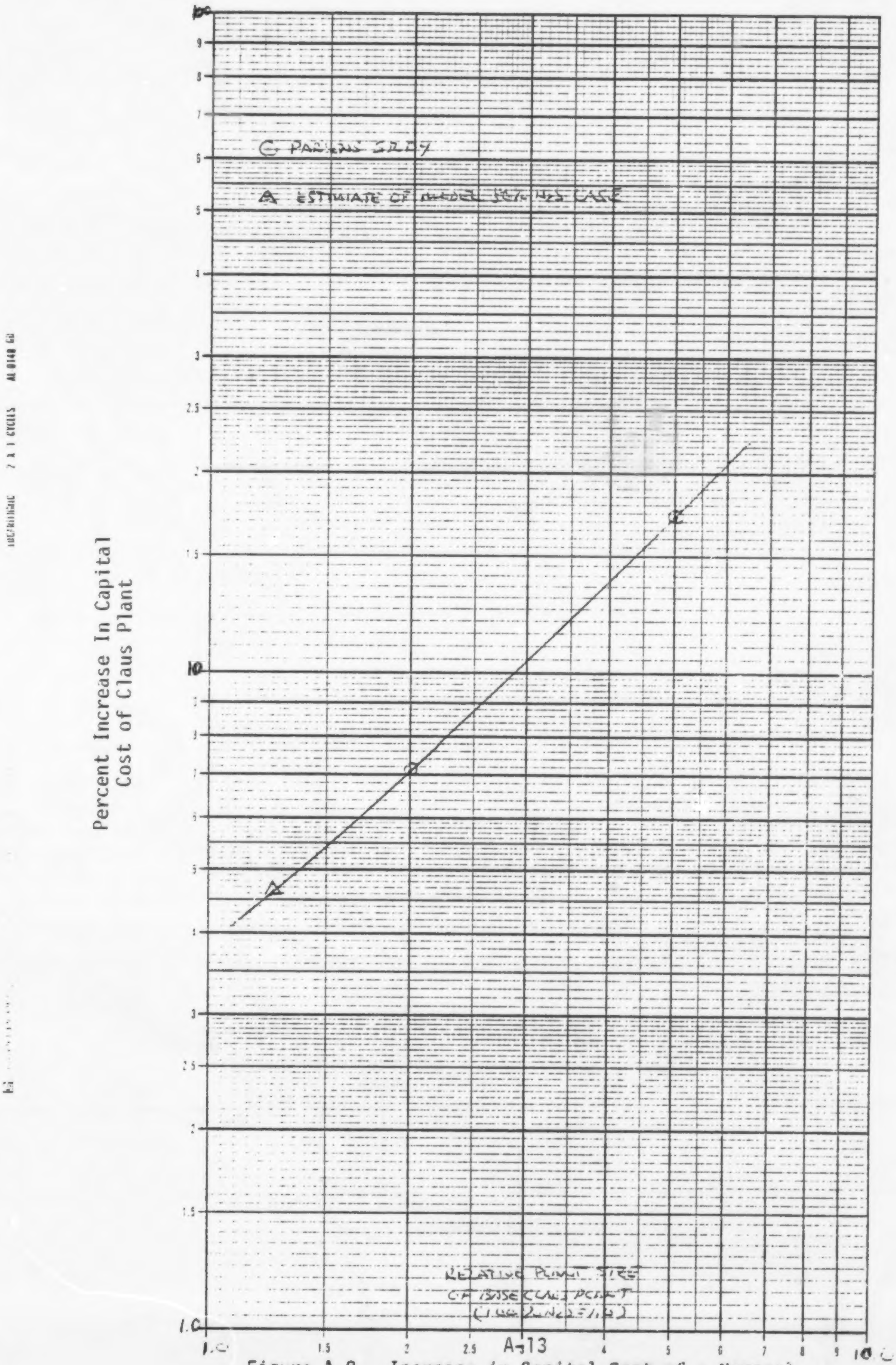


Figure A-8. Increase in Capital Cost of a Nominal 100 LT/D Claus Plant Due to Additional Tail Gas Treater

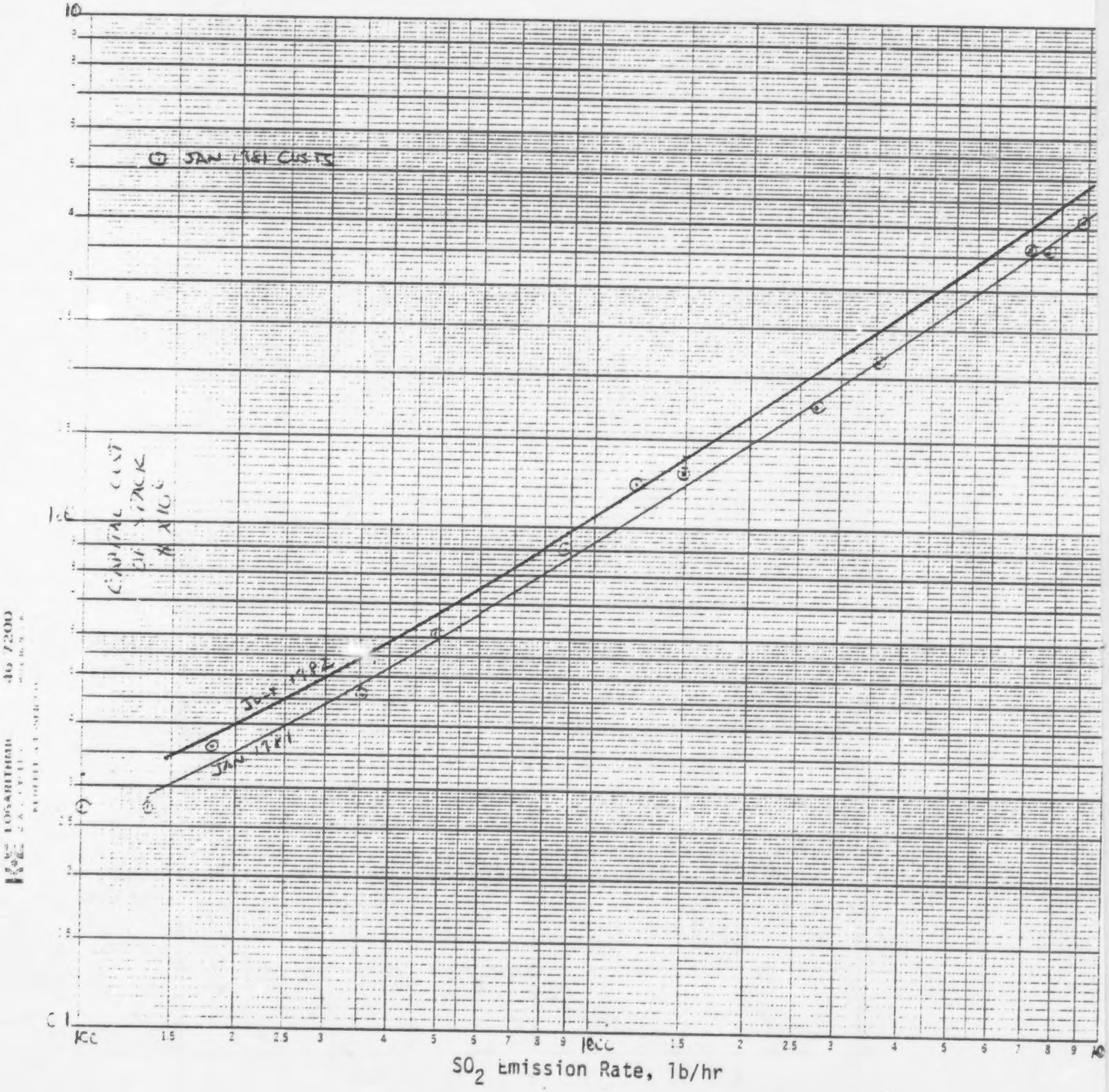


Figure A-9. Capital Cost of Stack vs. SO_2 Emission Rate

A.2. OPERATING COST ESTIMATES

In general, the operating costs were structured according to the methodology presented in the January 1983 draft Background Information Document for the Natural Gas Production Industry (EPA 450/3-82-023c).[7] Operating costs are broken down into the following categories:

- utility consumption and credits
- chemical consumption and credits
- labor-operating and supervisory
- maintenance and repair
- miscellaneous (supplies and laboratory changes)
- fixed costs - capital charges, taxes, and insurance
- overhead, including administrative and marketing

In lieu of actual cost data for refinery sulfur plant operations, the following costs and/or assumptions were extracted directly from the gas production document:

- utility prices and credits (see Table A-20)
- operating supplies and laboratory charges at 10 percent each of operating labor charges
- taxes - 1 percent of fixed capital costs
- insurance - 0.6 percent of fixed capital costs
- overhead - 25 percent of operating labor and maintenance
- administrative and marketing - 1 percent each of total annualized costs

Other operating cost estimates require more detailed explanation as in the following sections.

A.2.1. Utility Consumption and Credits

A.2.1.1 Claus Plants

Steam, feedwater, and electric power figures for Claus plants were estimated using graphs prepared from the Parsons study cited earlier. Figures A-10 and A-11 graphically illustrate steam and condensate production (consumption for 600 psig steam) in lbs/hr per long ton sulfur production as a function of gas flow for 2-stage and 3-stage Claus plants with no heat recovery; Figure A-12 shows similar figures for a heat recovery system as proposed by Parsons, based on incineration at 1200°F. Tables A-3, A-4, and A-5 show these data numerically for the three cases examined by

Figure A-10

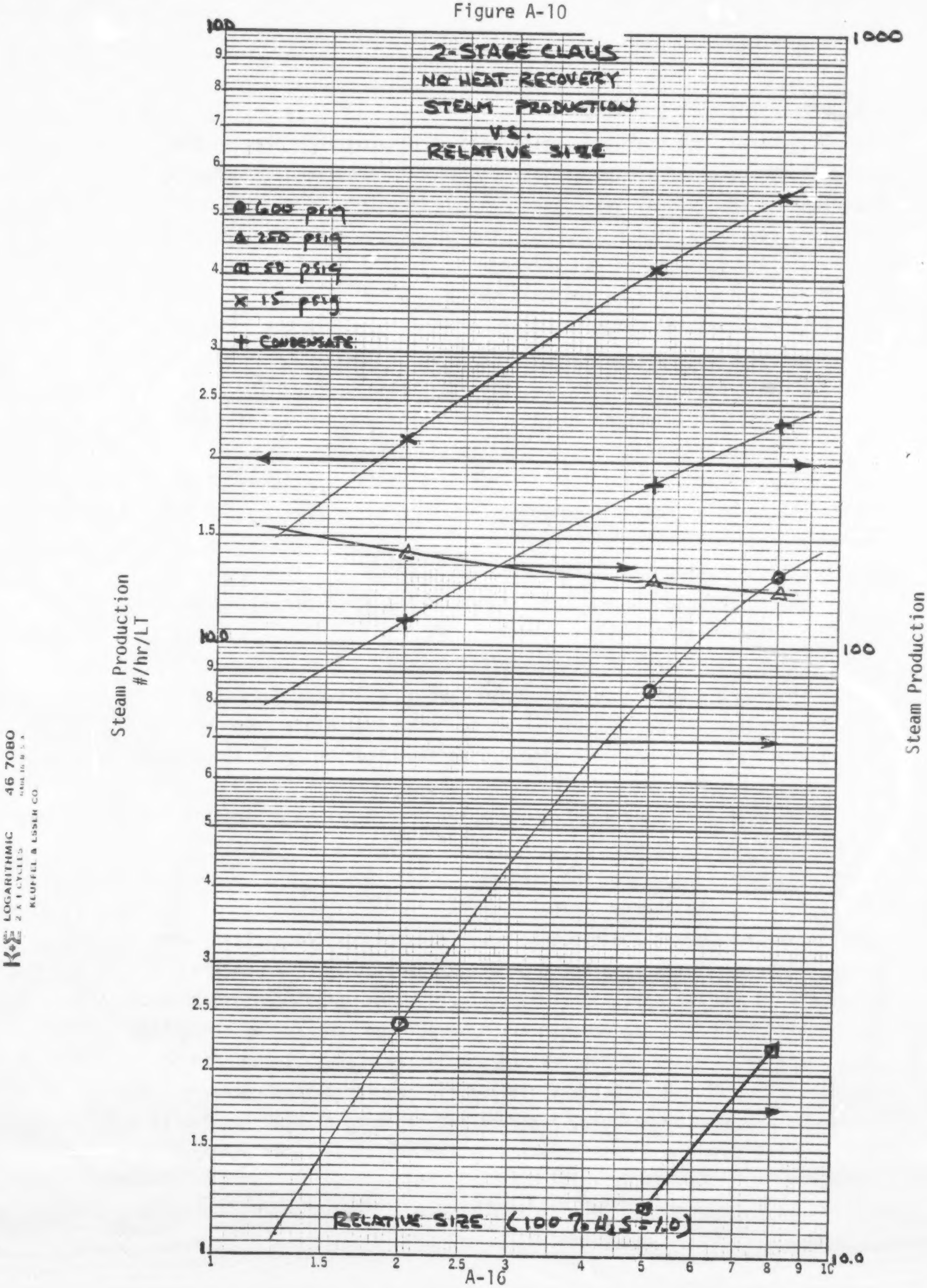
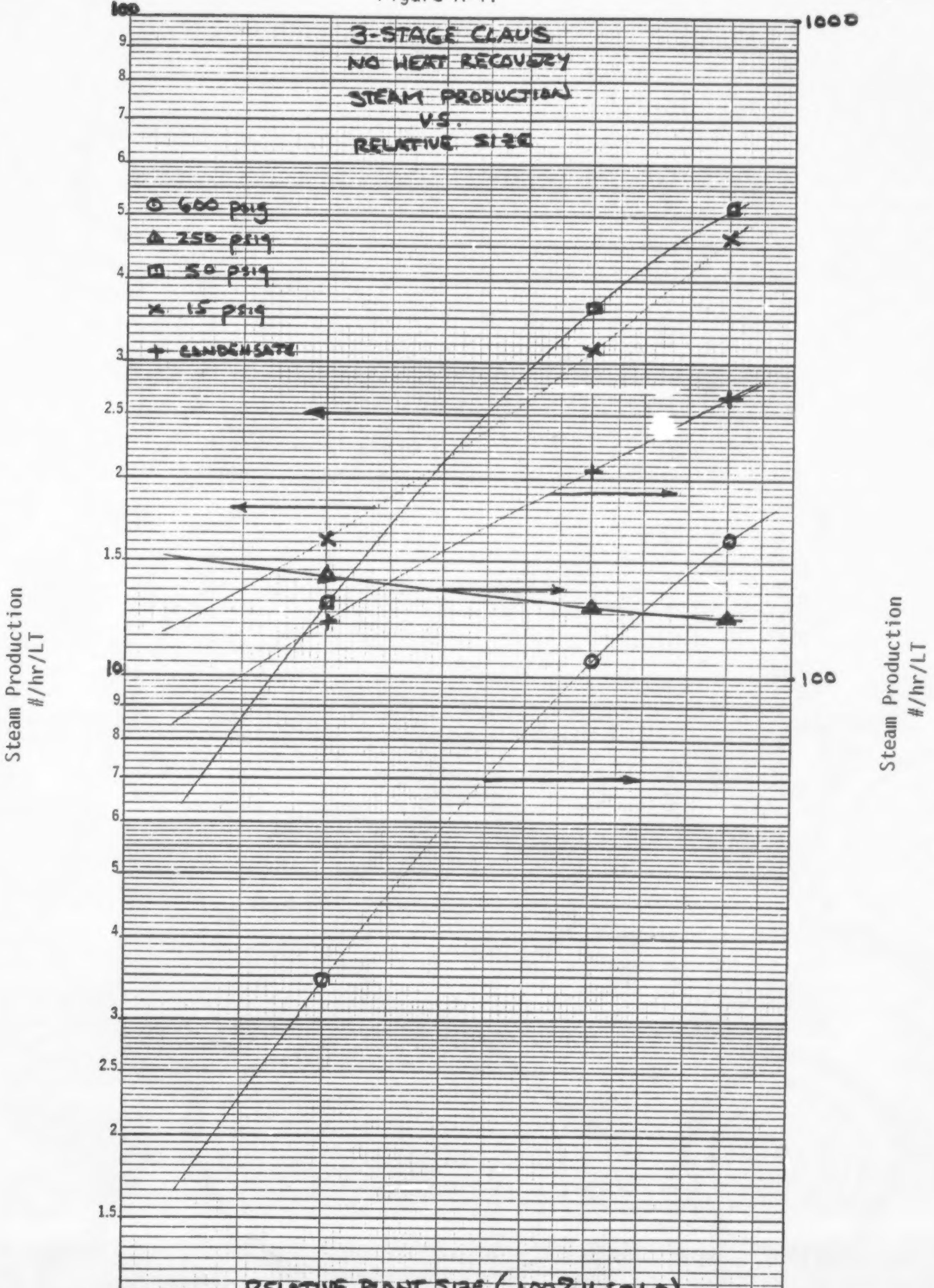

2-STAGE CLAUS
NO HEAT RECOVERY
STEAM PRODUCTION
VS.
RELATIVE SIZE
⊕ 600 psig
△ 250 psig
⊡ 50 psig
× 15 psig
+ CONDENSATE
Steam Production #/hr/LT
Steam Production #/hr/LT
RELATIVE SIZE (100 % H_2S = 1.0)
100
9
8
7
6
5
4
3
2.5
2
1.5
10.0
9
8
7
6
5
4
3
2.5
2
1.5
1
1000
100
10.0
1
1.5
2
2.5
3
4
5
6
7
8
9
10
K+E LOGARITHMIC 46 7080
2 X 1 CYCLES
KEUFFEL & ESSER CO.

Figure A-11



K+E LOGARITHMIC 2 X 1 CYCLES KEUFFEL & ESSER CO. 46 7080 MADE IN U.S.A.

Table A-3. 2-STAGE CLAUS
NO HEAT RECOVERY
STEAM PRODUCTION (#/HR/LT)

H_2S/CO_2 ratio	50/50	20/80	12.5/87.5	80/20 (estimated)
600 psig	(23.9)	(84.8)	(130.2)	(10.5)
250 psig	140.9	127.7	123.9	154
50 psig	0	12.2	22.1	0
15 psig	21.7	41.1	54.5	15.0
condensate	109.6	184.8	230.2	81.5

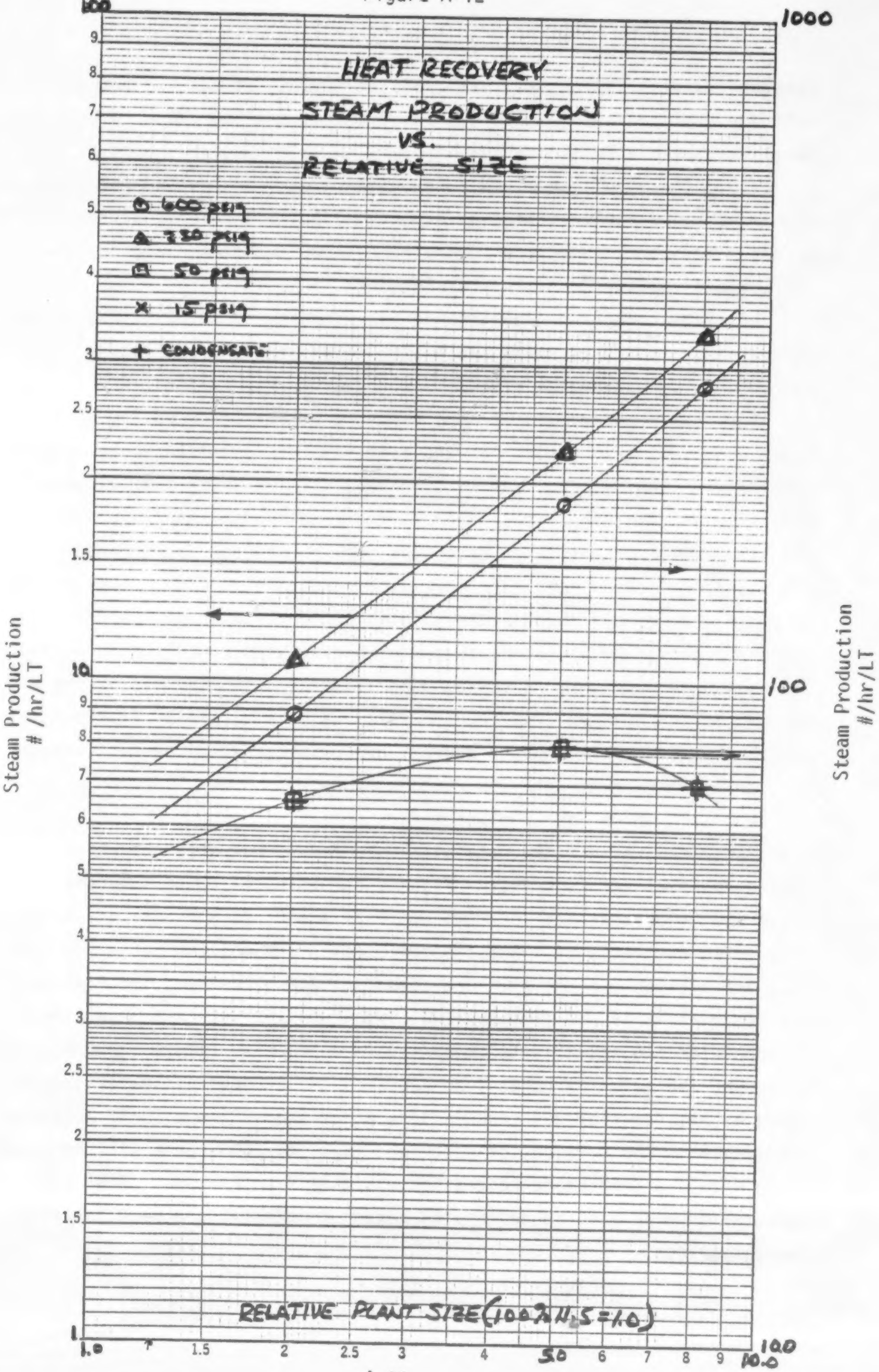
Table A-4. 3-STAGE CLAUS
NO HEAT RECOVERY
STEAM PRODUCTION (#/HR/LT)

H_2S/CO_2 ratio	50/50	20/80	12.5/87.5	80/20 (estimated)
600 psig	(34.3)	(105.7)	(161.5)	(17.5)
250 psig	140.9	127.7	123.9	150
50 psig	12.9	36.4	51.2	6.6
15 psig	26.2	31.4	46.1	12.4
condensate	120.2	205.7	261.5	88.0

Table A-5. CLAUS PLANT
HEAT RECOVERY
STEAM PRODUCTION (#/HR/LT)

H_2S/CO_2 ratio	50/50	20/80	12.5/87.5	80/20 (estimated)
600 psig	89.2	185.0	278.6	63.5
250 psig	10.7	22.14	33.46	7.4
50 psig	65.9	80.0	70.0	53.8
15 psig	--	--	--	--
condensate	65.9	80.0	70.0	53.8

Figure A-12



K+E LOGARITHMIC 46 7080
2 X 1 CYCLES MADE IN U.S.A.
KEUFFEL & ESSER CO.

Parsons and also include the extrapolated figures at 80% H_2S for a typical refinery application. Tables A-6 and A-7 then combine these results for model 2-stage and 3-stage Claus plants with heat recovery.

Using the total steam and condensate values, the boiler feed water requirements may be estimated by assuming a 2.7-3.0 percent system loss of steam and condensate; i.e., the total steam and condensate divided by .9715 equals boiler feedwater requirements.

Electric power requirements may be estimated by using either of two curves shown in Figures A-13 and A-14. These show electric power consumption as a function of gas flow and of sulfur in feed.

Fuel gas requirements for incinerators were calculated for each case based upon tail gas composition and temperatures according to principles outlined in Chemical Engineering Thermodynamics by Smith and Van Ness.[8] The calculations scheme is similar to that employed in Appendix C-II of EPA 450/2-78-012, Control of Emissions from Lurgi Coal Gasification Plants; page C-19 of that report is reprinted here as Figure A-15.[9] The only difference here involves recalculation of the average specific heats to correspond with the temperature ranges evaluated in this study--1200°F combustion temperature. Also fuel was assumed to be fuel gas having a composition of $C_{1.15}H_{4.3}$ having a heating value of 3.85×10^5 Btu/lb-mole (995.6 Btu/scf). All exhaust streams are oxidized at 25 percent excess air, to be consistent with the Parsons study.

A.2.1.2 Amine Tail Gas Treaters

There are very little data available for actual steam, electric power, and fuel gas consumption figures for amine tail gas treaters, since most reported data are combined with the Claus and fuel gas amine data. Two estimates of amine treater utility consumption for a 100 LT/D case are available along with one report of actual consumption figures for two systems of 170 LT/D and 240 LT/D.[10,11,12] Table A-8 shows these figures with the actual data converted to a 100 LT/D figure for comparison. As shown, the actual figures from ARCO and the Parsons estimates generally agree except for fuel gas consumption, where the ARCO and SOHIO estimates are similar. For purposes of model analyses, the ARCO data will be used, along with the condensate generation estimate from Parsons. Fuel gas consumption will also be calculated based on material and heat balances for comparison.

Table A-6. 2-STAGE CLAUS
WITH HEAT RECOVERY
STEAM PRODUCTION (#/HR/LT)

H_2S/CO_2 ratio	50/50	20/80	12.5/87.5	80/20 (estimated)
600 psig	65.3	100.2	148.4	53.0
250 psig	151.6	149.84	157.36	161.4
50 psig	65.9	92.2	92.1	53.8
15 psig	21.7	41.1	54.5	15.0
condensate	43.7	104.8	160.2	28.2

Table A-7. 3-STAGE CLAUS
WITH HEAT RECOVERY
STEAM PRODUCTION (#/HR/LT)

H_2S/CO_2 ratio	50/50	20/80	12.5/87.5	80/20 (estimated)
600 psig	54.9	79.3	117.1	46.0
250 psig	151.6	149.84	157.36	157.5
50 psig	78.8	116.4	121.2	60.4
15 psig	16.2	31.4	46.1	12.4
condensate	54.3	125.7	191.5	34.2

Figure A-13

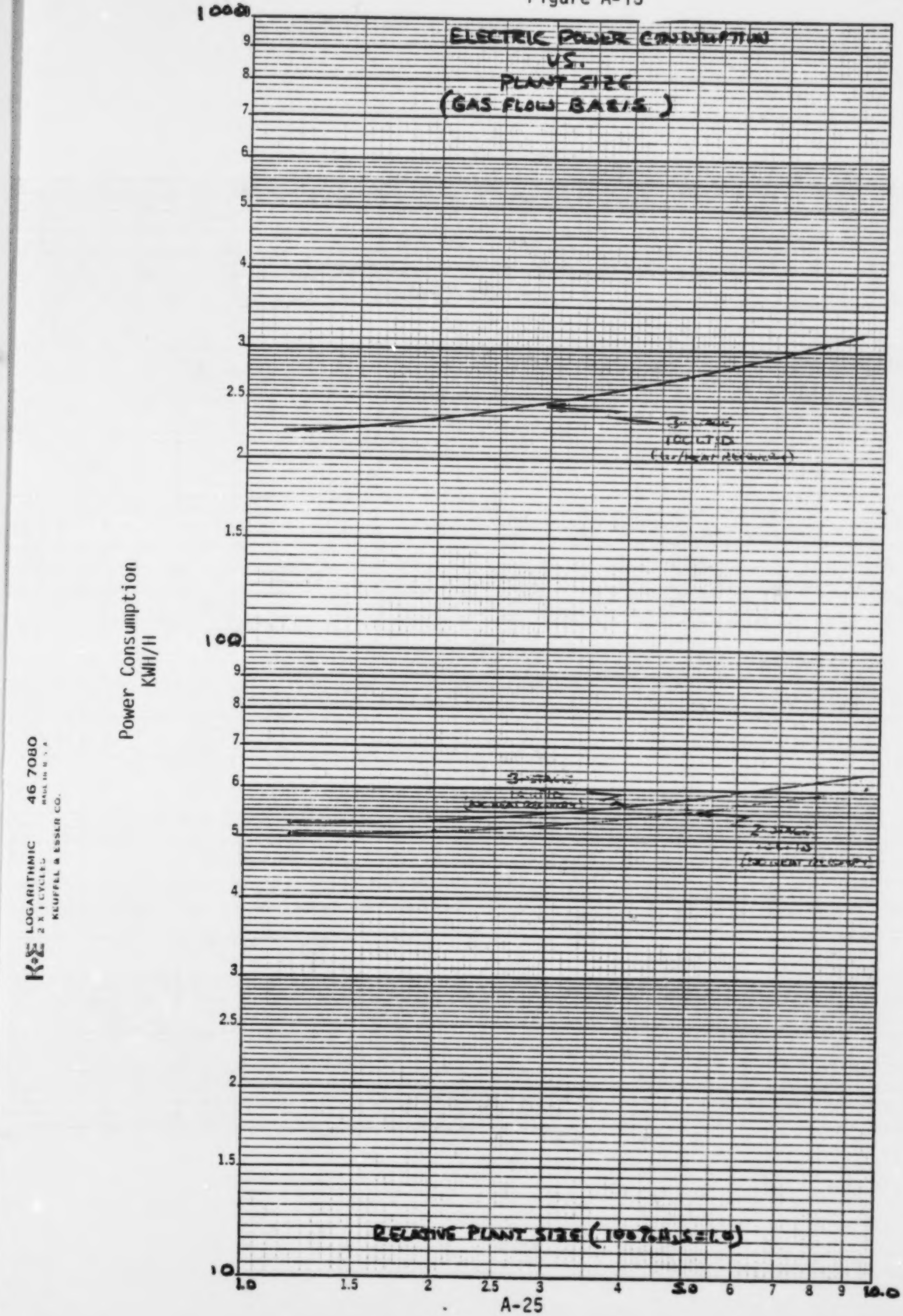
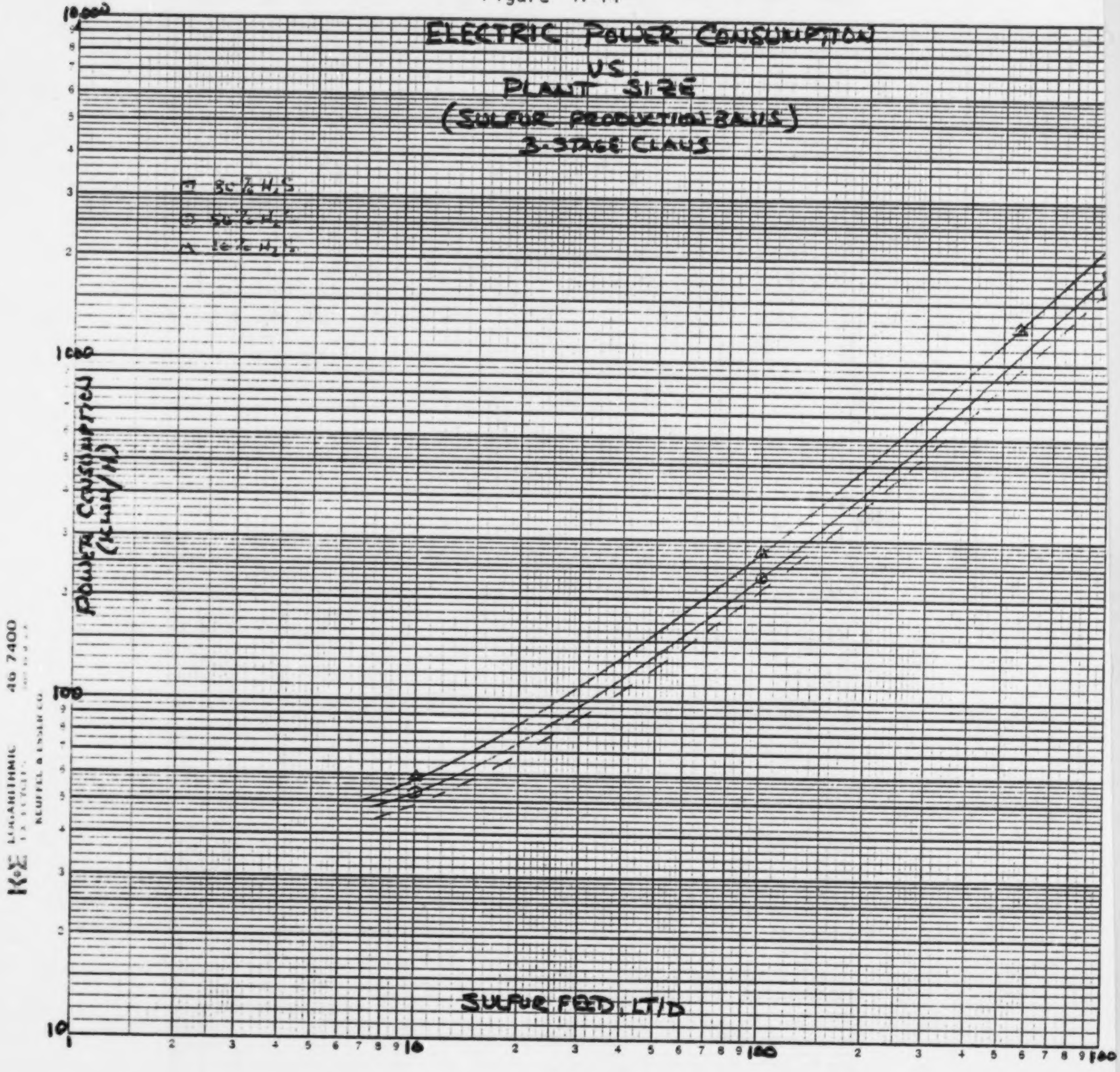

ELECTRIC POWER CONSUMPTION
VS.
PLANT SIZE
(GAS FLOW BASIS)
Power Consumption
KWH/H
1000
9
8
7
6
5
4
3
2.5
2
1.5
100
9
8
7
6
5
4
3
2.5
2
1.5
10
3-STAGE,
[illegible]
3-STAGE
[illegible]
2-STAGE
[illegible]
RELATIVE PLANT SIZE (100 T/H, S=1.0)
1.0
1.5
2
2.5
3
4
5.0
6
7
8
9
10.0
K+E LOGARITHMIC 2 X 1 CYCLES KEUFFEL & ESSER CO.
46 7080 MADE IN U.S.A.

Figure A-14


ELECTRIC POWER CONSUMPTION
VS.
PLANT SIZE
(SULFUR PRODUCTION BASIS)
3-STAGE CLAUS
□ 80% H_2S
○ 50% H_2S
△ 10% H_2S
POWER CONSUMPTION (KWH/H)
SULFUR FEED, LT/D
10,000
1000
100
10
1
10
100
1000
K+E LOGARITHMIC 46 7400
KEUFFEL & ESSER CO.

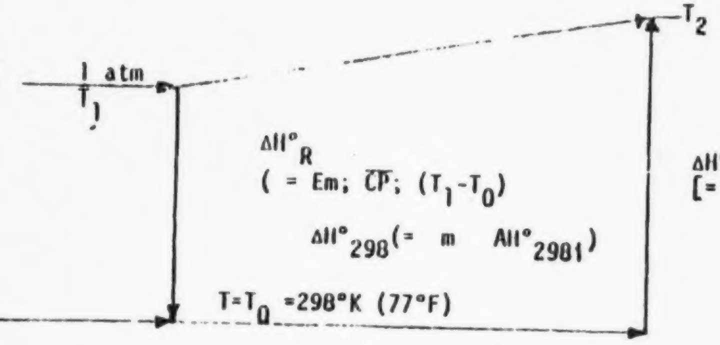
Reactants:

Claus tail gas @ 284°F
Stretford tail gas @ 77°F
Vent exp. gases @ 77°F

Fuel:

CH_4+
O_2
N_2

P_f = 1 atm
$T_f = T_o$ = 77°F



1 atm
T_1
T_2
$\Delta H°_R$
$(= \Sigma m_i\ \overline{C_P}_i\ (T_1 - T_0)$
$\Delta H°_{298}(= m\ \Delta H°_{298i})$
$T = T_0$ = 298°K (77°F)
$\Delta H°_P$
$[= m\ C_{pi}(T_2 - T_0)]$

Oxidized Products T_2=1600°F or 1144°K

$$C_{pi} = a_i + \frac{B_i}{2}(T+T_0) + \frac{X_i}{3}(T^2+TT_0+T_0^2)$$

Reactions:

1. $CO + 1/2\ O_2 \rightarrow CO_2$ $\Delta H°_{298}$ = 121,744.8 Btu/lb-mole
2. $H_2 + 1/2\ O_2 \rightarrow H_2O$ (g) $\Delta H°_{298}$ = 104,026.4 "
3. $CH_4 + 2O_2 \rightarrow CO_2 + 2H_6O$ (g) $\Delta H°_{298}$ = 345,135,6 "
4. $C_2H_4 + 3O_2 \rightarrow 2\ CO_2 + 2\ H_2O$ (g) $\Delta H°_{298}$ = 569,109.6 "
5. $C_2H_6 + 7/2\ O_2 \rightarrow 2\ CO_2 + 3\ H_2O$ (g) $\Delta H°_{298}$ = 614,309.4 "
6. $H_2S + 3/2\ O_2 \rightarrow H_2O$ (g) + SO_2 $\Delta H°_{298}$ = 233,097.4 "
7. $COS + 3/2\ O_2 \rightarrow CO_2 + SO_2$ $\Delta H°_{298}$ = 237,981.6 "
8. $C_3H_8 + 5O_2 \rightarrow 3\ CO_2 + 4\ H_2O$ (g) $\Delta H°_{298}$ = 955.080.0 "
9. $C_4H_{10} + 13/2\ O_2 \rightarrow 4\ CO_2 + 5\ H_2O$ (g) $\Delta H°_{298}$ =1237,752.0 "
10. $C_3H_6 + 9/2\ O_2 \rightarrow 3\ CO_2 + 3\ H_2O$ (g) $\Delta H°_{298}$ = 885,582.0 "
11. $C_4H_8 + 6O_2 \rightarrow 4\ CO_2 + 4\ H_2O$ (g) $\Delta H°_{298}$ =1169,010.0 "
12. $CH_3OH + 3/2\ O_2 \rightarrow CO_2 + 2\ H_2O$ (g) $\Delta H°_{298}$ = 290,786.4 "
13. $S + O_2 \rightarrow SO_2$ $\Delta H°_{298}$ = 127,728.0 "

Component	$\overline{Cp}$(T=284°F)	$\overline{Cp}$(T=1600°F)
CO	6.99	-
H_2	6.94	-
CH_4	9.25	-
C_2H_4	11.89	-
C_2H_6	14.42	-
N_2	6.97	8.14
CO_2	9.29	13.04
H_2S	8.38	-
COS	10.62	-
SO_2	11.76	12.87
H_2O	8.11	10.77
O_2	-	8.44
CS_2	11.95	-
S	10.82	-

Air/Fuel Requirements: Required CH_4 = x lb-moles/hr

$O_2 = 1.2[1/2\ (CO) + 1/2\ (H_2) + 2\ (CH_4) + 3\ (C_2H_4) + 7/2\ (C_2H_6) + 5\ (C_3H_8) + 13/2\ (C_4H_{10}) + 9/2\ (C_3H_6) + 6\ (C_4H_8) + 3/2\ (CH_3OH) + 3/2\ (H_2S + COS)]$

$N_2 = 79/21\ O_2$

Assuming fuel gas is 970 BTU/SCF as CH_4,

$$x = \frac{\Delta H°_P - \Delta H°_{298} - \Delta H°_R}{(970\ \text{BTU/SCF})(386.7\ \text{SCF/lb-mole})} = \frac{\Delta H°_P - \Delta H°_{298} - \Delta H°_R}{3.75 \times 10^5\ \text{BTU/lb-mole}}$$

Figure A-15. Incineration Calculations

Table A-8. UTILITY CONSUMPTION BY AMINE TAIL GAS UNITS
(100 LT/D Claus basis)

Source	Power, kw	Cooling water, GPM	50# steam lb/hr	Fuel gas 10^6 Btu/hr	Condensate Generated lb/hr
SOHIO (estimate)	180	170	8390	5.33	---
ARCO (actual)	175	870	12472	5.2	---
PARSONS (estimate)	200	840	13277	6.5	14,300

A.2.2. Catalyst and Chemical Consumption

Catalyst consumption figures for Claus plants are based on an assumption that the first stage catalyst is replaced on a two-year cycle, the second stage at four years, and the third at six years. The assumed catalyst is alumina at $17 per cubic foot or $765 per short ton ($856.80 per long ton). Catalyst charge is estimated at 230 pounds per reactor per long ton Claus capacity for 80 percent H_2S feed.[13,14]

For amine tail gas units, catalyst replacement (cobalt-molybdenum) for the reduction reactor is assumed to be once every two years. The catalyst charge is assumed to be about 1/2 of a Claus stage, or 115 pounds per reactor per long ton Claus capacity (80% H_2S feed). The assumed catalyst cost is 10 times that of the Claus catalyst or $8568/long ton.[15]

Tail gas chemical consumption is a more elusive subject as amine type, degree of fouling, and degree of enhanced recovery by use of defoaming agents and organic contaminant removal varies from plant to plant. Actual figures provided for three systems show consumption of DIPA at 0.56, 0.67, and 3.1 lb/hr per 100 LT/D parent Claus capacity, averaging 1.44 lb/hr per 100 LT/D.[16,17,18]

For model purposes, a figure of 0.70 lb/hr per 100 LT/D Claus capacity will be used.

A.2.3. Labor, Maintenance, and Repair Costs

In the Parsons study, labor costs were estimated as follows--Claus plants @ 1.25 operators per shift, Claus + tail gas treater @ 2.25 operators per shift. Supervision was assumed at 0.25 per shift for both cases.

From two new operating plants having both Claus plants and tail gas treatment, the following data were obtained:[19,20]

Plant	Start-up Date	Claus Capacity	Type of Tail Gas Treater	Manpower/shift	
				Claus	Tail Gas
1	1980	100	Amine	2/3	2/3
2	1981	475	Amine	2/3	2/3

Hence, for new plants the labor assumptions are 2/3 operator per shift each for the Claus and tail gas treater and 1/4 supervisor per shift. Hourly rates per the Gas Production NSPS study are \$14.50/hr for operators and \$18.80/hr for supervision.

Maintenance and repair costs also varied widely from plant to plant. For Claus plants, labor and materials ranged from 2.3 to 6.1 percent of estimated fixed capital costs for Claus plants and 2.1 to 6.3 percent of capital costs for tail gas units. Other studies have assumed 3 percent of fixed capital (EPA 1975) and 3.5 of fixed capital (Gas Production NSPS Document, 1983).

Since the average labor and materials cost from the six new NSPS and two dozen or so older units was about 3 percent for both systems, this figure is assumed for model purposes.

A.3. MODEL PLANT LINE COSTS

A.3.1. Capital Cost and Operating Parameter Estimates

Using the economic assumptions and cost curves presented in the first two sections of this Appendix, the following model plants were evaluated:

Case	Claus Plant, LT/D	Tail Gas Treater
1A	2 stage, 10 LT/D	
1B	2 stage, 10.48 LT/D	Amine 0.96 LT/D design
2A	3 stage, 50 LT/D	
2B	3 stage, 51.65 LT/D	Amine 3.3 LT/D design
3A*	3 stage, 100 LT/D	
3B*	3 stage, 103.3 LT/D	Amine 6.6 LT/D design

* Waste heat boiler included for incinerator

Case 1A

Key line item estimates for case 1A are:

Item	Source	Estimate (July 1982)
Capital Cost:		
2-stage Claus	Figure A-2	$\$1.97 \times 10^6$
Incinerator	Figure A-4	$\$0.26 \times 10^6$
Stack	Figure A-9	$\$0.31 \times 10^6$
		$\$2.54 \times 10^6$

Item	Source	Estimate (July 1982)
Operating Cost (Credit):		
600 psig steam	Figure A-10, Table A-3	105 #/hr
250 psig steam	" "	(1540 #/hr)
15 psig steam	" "	(150 #/hr)
condensate	" "	(~~81.5~~ #/hr) [illegible]
electric power	Figure A-13	50.5 KWH/hr
fuel gas	Figure A-15 (calculated)	0.60 10^6 Btu/hr
catalyst	Section A.2.2	1725 #/yr

Case 1B 10.48 LT/D (78.07% H_2S)

Key line item estimates for case 1 B are:

Item	Source	Estimate: Claus	Estimate: Tail Gas
Capital Cost:			
2-stage Claus	Equation A-1 + Figure A-2	$\$2.04 \times 10^6$	
Amine Treater	Figure A-5		$\$2.35 \times 10^6$
Incinerator	Figure A-4	$\$0.26 \times 10^6$	
Stack	Figure A-9	$\$0.31 \times 10^6$	
		$\$2.61 \times 10^6$	
Operating Cost:			
600 psig	Figure A-10	117 #/hr	--
250 psig	"	(1603 #/hr)	--
50 psig steam	Table A-8		1814 #/hr
15 psig steam	Figure A-10	(160 #/hr)	
15 condensate	Figure A-10/	(867 #/hr)	(2080 #/hr)
electric power	Figure A-13/ Table A-8	51 KWH/H	39.5 KWH/H
fuel gas	Figure A-15 (calculated)	0.58×10^6 Btu/hr	0.76×10^6 Btu/hr
cooling water		--	126.5 gpm
catalyst	Section A.2.2	1852 #/yr	617.5 #/yr
chemicals:			
DIPA	Section A.2.2	--	616 lb/yr
Soda	"	--	1000 lb/yr

Case 2A 50 LT/D (80% H_2S)

Item	Source	Estimate
Capital Cost:		
3-stage Claus	Figure A-2	$\$3.50 \times 10^6$
Incinerator	Figure A-4	0.38×10^6
Stack	Figure A-9	0.51×10^6
		$\$4.33 \times 10^6$

Item	Source	Estimate
Operating Cost (Credit):		
600 psig steam	Table A-4	875 #/hr
250 psig steam	"	(7500 #/hr)
50 psig steam	"	(330 #/hr)
15 psig steam	"	(620 #/hr)
condensate	"	(4400 #/hr)
electric power	Figure A-14	125 KWH/H
fuel gas	Figure A-15 (calculated)	3.0×10^6 Btu/hr
catalyst	Section A.2.2	10,542 lb/yr

Case 2B 51.65 LT/D (78.7% H_2S)

Item	Source	Estimate	
		Claus	Tail Gas
Capital Cost:			
3-stage Claus	Equation A-1 + Figure A-2	$\$3.60 \times 10^6$	
Amine Treater	Figure A-6		$\$3.60 \times 10^6$
Incinerator	Figure A-4	$\$0.32 \times 10^6$	
Stack	Figure A-9	$\$0.31 \times 10^6$	
		$\$4.23 \times 10^6$	
Operating Cost:			
600 psig steeam	Figure A-11	919 #/hr	--
250 psig steam	"	(7748 #/hr)	--
50 psig steam	" /Table A-8	(346 #/hr)	623 lb/hr
15 psig steam	"	(635 #/hr)	--
condensate	" /Table A-8	(4597 #/hr)	(7150 lb/hr)
electric power	Figure A-14/Table A-8	128 KWH/H	99 KWH/H
fuel gas	Figure A-15 (calculated)	209×10^6 Btu/hr	2.6×10^6 Btu/hr
cooling water	Table A-8		435 gpm
catalyst	Section A.2.2	11,067 lb/yr	3180 lb/yr
chemicals:			
DIPA	"		3037 lb/yr
Soda	"		4928 lb/yr

Case 3A 100 LT/D (80% H_2S)

Item	Source	Estimate
Capital Cost:		
3-stage Claus	Figure A-2	$\$4.50 \times 10^6$
Incinerator	Figure A-4	0.41×10^6
Stack	Figure A-9	0.75×10^6
Waste Heat Recovery System	Figure A-5	0.56×10^6
		$\$6.26 \times 10^6$
Operating Cost (Credit):		
600 psig	Table A-7	(4600 lb/hr)
250 psig	"	(15740 lb/hr)
50 psig	"	(6040 lb/hr)
15 psig	"	(1240 lb/hr)
condensate	"	(3420 lb/hr)
electric power	Figure A-14	212 KWH/H
fuel gas		6.0×10^6 Btu/hr
catalyst	Section A.2.2	21,084 lb/yr

Case 3B 103.3 LT/D (78.7% H_2S)

Item	Source	Estimate	
		Claus	Tail Gas
Capital Costs:			
3-stage Claus	Equation A-1 + Fig. A-2	$\$4.63 \times 10^6$	
Amine Treater	Figure A-6		$\$4.68 \times 10^6$
Incinerator	Figure A-4	0.41×10^6	
Stack	Figure A-9	0.31×10^6	
Waste Heat Recovery System	Figure A-5	0.57×10^6	
		$\$5.92 \times 10^6$	
Operating Cost (Credit):			
600 psig	Figure A-11	(4750 lb/hr)	
250 psig	"	(16260 lb/hr)	
50 psig	", Table A-8	(6240 lb/hr)	12,472 lb/hr
15 psig	Figure A-11	(1280 lb/hr)	
condensate	", Table A-8	(3530 lb/hr)	(14300 lb/hr)
electric power	Figure A-14/Table A-8	215 KWH/H	165 KWH/H
fuel gas	calculated	7.15×10^6 Btu/hr	5.2×10^6 Btu/hr
cooling water	Table A-8		870 gpm
catalyst	Section A.2.2	22,134 lb/yr	6360 lb/yr
chemicals:			
DIPA	"		6074 lb/yr
Soda	"		9856 lb/yr

Combining the above figures with the prices in Table A-9 results in line item costs as presented in Table A-10. A significant portion of annual operating costs is the capital recovery factor. For comparison, Table A-11 shows the annual costs, and costs per ton SO_2 controlled for interest rates of 10, 15, and 20 percent for a 15-year lifetime.

Table A-9. ECONOMIC ASSUMPTIONS USED TO CALCULATE ANNUALIZED COSTS[a]

I. Utility prices:

1.	600 psig steam	\$15.98/Mg	(\$7.25/1,000 lb)
2.	250 psig steam	\$14.88/Mg	(\$6.75/1,000 lb)
3.	50 psig steam	\$12.68/Mg	(\$5.75/1,000 lb)
4.	15 psig steam	\$ 9.92/Mg	(\$4.50/1,000 lb)
5.	boiler feedwater	\$ 3.31/Mg	(\$1.50/1,000 lb)
6.	steam condensate	\$ 2.76/Mg	(\$1.25/1,000 lb)
7.	cooling water	\$13.21/10^{3}m^3	(\$ .05/1,000 gal)
8.	catalyst:		
	a. alumina	\$352.64/Mg	(\$0.38/lb)[c]
	b. cobalt-molybdenum (Co/Mo)	\$3,5256/Mg	(\$3.80/lb)[c]
9.	Chemicals:		
	a. diisopropanolamine	\$0.49/Kg	(\$1.07/lb)[b]
	b. soda	\$330.6/Mg	(\$300/ton)[b]
10.	fuel gas	\$3.64/10^9/J	(\$3.50/10^6 Btu)[d]
11.	electric power	\$0.05/KWH	
12.	sulfur	\$118.08 Mg	(\$120/LT)[e]

II. Labor (8,760 hours per year basis)

1. operators: (\$14.50/hr)
   2/3 per shift for Claus
   2/3 per shift for tail gas treater

2. supervision: (\$18.80/hr)
   1/4 per shift for sulfur recovery facility

III. Maintenance and Repair

Labor and materials: 3.0 percent of fixed capital Costs[e]

IV. Other Miscellaneous Costs

1. Operating supplies: 10 percent of operating labor
2. Laboratory charges: 10 percent of operating labor

V. Fixed Charges

1. Capital charges = fixed capital costs x $\frac{i(1+i)^n}{(1+i)^n-1}$
   = a) .13148 for n = 15 years and i = 10%
   b) .17106 for n = 15 years and i = 15%
   c) .21382 for n = 15 years and i = 20%

2. Local taxes - 1 percent of fixed capital costs

3. Insurance - 0.6 percent of fixed capital costs

Table A-9. ECONOMIC ASSUMPTIONS USED TO CALCULATE ANNUALIZED COSTS[a] (continued)

VI. Overhead

1. plant overhead - 25 percent of operating labor + 25 percent of maintenance and repair
2. administrative - 1 percent of annualized costs
3. distribution and marketing - 1 percent of annualized costs

---

[a] All assumptions and values assigned from Reference 1 unless otherwise noted; actual consumption figures for model plants from EPA survey and Reference 2.

[b] Chemical Market Reporter, October 4, 1982.

[c] Telephone conversation with Mr. R. E. Warner of Ralph M. Parsons Co., February 1, 1983.

[d] Memorandum: R. E. Jenkins to C. B. Sedman, EPA, dated September 7, 1982.

[e] Average of EPA survey.

Table A-10. LINE ITEM COSTS FOR MODEL PLANTS

MODEL 1A (10.16 Mg/d)

Capital cost - $\$2.54 \times 10^6$

	i = 15%	i = 10%	i = 20%
Direct operating cost			
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 6,395	$ 6,395	$ 6,395
2. treated boiler feedwater	21,615	21,615	21,615
3. electric power	21,210	21,210	21,210
4. fuel gas	17,640	17,640	17,640
5. catalyst	655	655	655
B. Labor			
1. Operators	$84,680	$84,680	$84,680
2. Supervision	41,170	41,170	41,170
C. Maintenance and Repair	$76,200	$76,200	$76,200
D. Supplies and laboratory charges	$16,940	$16,940	$16,940
Fixed Charges:			
A. Capital	$434,490	$333,960	$543,105
B. Taxes	25,400	25,400	25,400
C. Insurance	15,240	15,240	15,240
Plant Overhead:	$40,220	$40,220	$40,220
General Expenses			
A. Administrative	$ 8,020	7,160	9,100
B. Distribution and sales	$ 8,020	7,160	9,100
Total Annualized Costs	$817,895	715,645	928,670
Credits			
1. 1,960 Kp steam	$ 87,320	$ 87,320	$ 87,320
2. 106 Kp steam	5,670	5,670	5,670
3. steam condensate	8,558	8,558	8,558
4. sulfur	399,420	399,420	399,420
Total Credits	$499,265	$499,265	$499,265
Net Annual Operating Cost for Case 1A	$320,439	$218,189	$431,214

Table A-10. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 1B (10.16 Mg/d)

Capital Cost - $4.96 x 10^6

Direct operating cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 7,125	$ 7,125	$ 7,125
2. 352 Kp steam	87,615	87,615	87,615
3. treated boiler feedwater	36,040	36,040	36,040
4. electric power	38,010	38,010	38,010
5. fuel gas/hydrogen	39,395	39,395	39,395
6. cooling water	3,190	3,190	3,190
7. catalyst	3,050	3,050	3,050
8. chemicals	810	810	810
B. Labor			
1. Operators	$169,360	169,360	169,360
2. Supervision	82,340	82,340	82,340
C. Maintenance & Repair	$148,800	$148,800	$148,800
D. Supplies & Lab Charges	$ 33,870	$ 33,870	$ 33,870
Fixed Charges			
A. Capital	$848,460	652,140	1,060,545
B. Taxes	49,600	49,600	49,600
C. Insurance	29,760	29,760	29,760
Plant Overhead:	$ 79,540	79,540	79,540
General Expenses			
A. Administrative	$ 16,550	14,650	18,750
B. Distribution and sales	16,550	14,650	18,750
Total Annualized Costs	$1,690,065	1,489,945	1,906,550
Credits			
1. 1,960 Kp steam	$ 90,890	90,890	90,890
2. 106 Kp steam	6,050	6,050	6,050
3. steam condensate	30,945	30,945	30,945
4. sulfur	419,580	419,580	419,580
Total Credits	$547,465	$547,465	$547,465
Net Annual Operating Cost for Case 1B	$1,142,600	$942,480	$1,359,085

Table A-10. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 2A (50.8 Mg/D)

Capital Cost - $4.33 x 10^6

Direct Operating Cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 53,290	$ 53,290	$ 53,290
2. treated boiler feedwater	155,310	155,310	155,310
3. electric power	52,500	52,500	52,500
4. fuel gas	88,200	88,200	88,200
5. catalyst	4,005	4,005	4,005
B. Labor			
1. Operators	84,680	84,680	84,680
2. Supervision	41,170	41,170	41,170
C. Maintenance and Repair	129,900	129,900	129,900
D. Supplies and Lab Charges	16,940	16,940	16,940
Fixed Charges			
A. Capital	740,690	569,310	925,840
B. Taxes	43,300	43,300	43,300
C. Insurance	25,980	25,980	25,980
Plant Overhead	53,645	53,645	53,645
General Expenses			
A. Administrative	15,000	13,300	16,850
B. Distribution and Sales	10,000	13,300	16,850
Total Annualized Costs	$1,519,610	1,344,830	1,708,460
Credits			
1. 1,960 Kp steam	$425,250	$425,250	$425,250
2. 352 Kp steam	15,940	15,940	15,940
3. 106 Kp steam	23,435	23,435	23,435
4. steam condensate	46,200	46,200	46,200
5. sulfur	2,028,600	2,028,600	2,028,600
Total Credits	$2,539,425	2,539,425	2,539,425
Net Annual Operating Cost for Case 2A	($1,019,815)	(1,194,595)	(830,965)

Table A-10. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 2B (50.8 Mg/D)

Capital Cost - $7.83 x 10^6

	i = 15%	i = 10%	i = 20%
Direct Operating Cost			
A. Utilities & Chemicals			
1. 4,300 Kp steam	$ 55,965	$ 55,965	$ 55,965
2. 352 Kp steam	284,485	284,485	284,485
3. treated boiler feedwater	172,770	172,770	172,770
4. electric power	95,340	95,340	95,340
5. fuel gas/hydrogen	161,000	161,000	161,000
6. cooling water	10,960	10,960	10,960
7. catalyst	16,290	16,290	16,290
8. chemicals	3,990	3,990	3,990
B. Labor			
1. Operators	169,360	169,360	169,360
2. Supervision	82,340	82,340	82,340
C. Maintenance & Repair	234,900	234,900	234,900
D. Supplies & Lab Charges	33,870	33,870	33,870
Fixed Charges			
A. Capital	1,339,400	1,029,490	1,674,210
B. Taxes	78,300	78,300	78,300
C. Insurance	46,980	46,980	46,980
Plant Overhead	101,065	101,065	101,065
General Expenses			
A. Administrative	27,750	24,650	31,100
B. Distribution & Sales	27,750	24,650	31,100
Total Annualized Cost	2,842,150	2,526,040	3,183,660
Credits			
1. 1,960 Kp steam	439,310	439,310	439,310
2. 106 Kp steam	24,005	24,005	24,005
3. steam condensate	123,345	123,345	123,345
4. sulfur	2,097,900	2,097,900	2,097,900
Total Credits	$2,684,560	$2,684,560	$2,684,560
Net Annual Operating Cost for Case 2B	$ 157,590	($158,520)	$ 499,100

Table A-10. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 3A (101.6 Mg/D)

Capital cost - $6.26 x 10^6

	i = 15%	i = 10%	i = 20%
Direct Operating Cost			
A. Utilities & Chemicals			
1. treated boiler feedwater	$402,575	$402,575	$402,575
2. electric power	89,040	89,040	89,040
3. fuel gas	176,400	176,400	176,400
4. catalyst	8,010	8,010	8,010
B. Labor			
1. Operators	84,680	84,680	84,680
2. Supervision	41,170	41,170	41,170
C. Maintenance & Repair	187,800	187,800	187,800
D. Supplies & Lab Charges	16,940	16,940	16,940
Fixed Charges			
A. Capital	1,070,835	823,065	1,338,515
B. Taxes	62,600	62,600	62,600
C. Insurance	37,560	37,560	37,560
Plant Overhead	68,120	68,120	68,120
General Expenses			
A. Administrative	22,460	19,980	25,135
B. Distribution & Sales	22,460	19,980	25,135
Total Annualized Costs	$2,290,650	2,037,890	2,563,680
Credits			
1. 4,300 Kp steam	280,140	280,140	280,140
2. 1,960 Kp steam	92,460	92,460	92,460
3. 352 Kp steam	291,730	291,730	291,730
4. 106 Kp steam	46,870	46,870	46,870
5. steam condensate	35,910	35,910	35,910
6. sulfur	4,057,200	4,057,200	4,057,200
Total Credits	$5,604,310	5,604,310	5,604,310
Net Annual Operating Cost for Case 3A	($3,313,660)	($3,566,420)	($3,040,630)

Table A-10. LINE ITEM COSTS FOR MODEL PLANTS (continued)

MODEL 3B (101.6 Mg/D)

Capital cost - $\$10.60 \times 10^6$

Direct Operating Cost	i = 15%	i = 10%	i = 20%
A. Utilities & Chemicals			
1. 352 Kp steam	$301,005	$301,005	$301,005
2. treated boiler feedwater	414,820	414,820	414,820
3. electric power	159,600	159,600	159,600
4. fuel gas/hydrogen	363,090	363,090	363,090
5. cooling water	21,925	21,925	21,925
6. catalyst	32,580	32,580	32,580
7. chemicals	7,980	7,980	7,980
B. Labor			
1. Operators	169,360	169,360	169,360
2. Supervision	82,340	82,340	82,340
C. Maintenance & Repair	318,000	318,000	318,000
D. Supplies & Lab Charges	33,870	33,870	33,870
Fixed Charges			
A. Capital	1,813,235	1,393,690	2,266,490
B. Taxes	106,000	106,000	106,000
C. Insurance	63,600	63,600	63,600
Plant Overhead	121,840	121,840	121,840
General Expenses			
A. Administrative	39,750	35,550	44,280
B. Distribution & Sales	39,750	35,550	44,280
Total Annualized Cost	$4,088,745	3,660,800	4,551,060
Credits			
1. 4,300 Kp steam	289,275	289,275	289,275
2. 1,960 Kp steam	921,940	921,940	921,940
3. 106 Kp steam	48,385	48,385	48,385
4. steam condensate	187,215	187,215	187,215
5. sulfur	4,195,800	4,195,800	4,195,800
Total Credits	$5,642,615	5,642,615	5,642,615
Net Annual Operating Cost for Case 3B	($1,553,870)	($1,981,815)	($1,091,555)

Table A-11. COST & COST-EFFECTIVENESS OF MODEL CONTROLS

	Plant Size, LT/D		
i = 10 percent	10	50	100
Base Case Annual Cost, $	218,189	($1,194,595)	($3,566,420)
Base Case SO_2 Removed, tons/yr	7,455.84	37,867.2	75,734.4
NSPS Case Annual Cost, $	$942,480	($158,520)	($1,981,815)
NSPS Case SO_2 Removed, tons/yr	7,832.16	39,160.8	78,321.6
Cost-Effectiveness, $/ton	$1,929	$801	$612
i = 15 percent			
Base Case Annual Cost, $	320,439	($1,019,815)	($3,313,660)
Base Case SO_2 Removed, tons/yr	7,455.84	37,867.2	75,734.4
NSPS Case Annual Cost, $	$1,142,600	$157,590	($1,553,870)
NSPS Case SO_2 Removed, tons/yr	7,832.16	39,160.8	78,321.6
Cost-Effectiveness, $/ton	$2,190	$910	$680
i = 20 percent			
Base Case Annual Cost, $	$431,214	($830,965)	($3,040,630)
Base Case SO_2 Removed, tons/yr	7,455.84	37,867.2	75,734.4
NSPS Case Annual cost, $	$1,359,085	$499,100	($1,091,555)
NSPS Case SO_2 Removed, tons/yr	7,832.16	39,160.8	78,321.6
Cost-Effectiveness, $/ton	$2,471	$1,028	$753

A.4 REFERENCES

1. Sulfur Recovery Study - Onshore Sour Gas Production Facilities, Job No. 6165-1, Ralph M. Parsons Company, July 1981.

2. Standards Support and Environmental Impact Statement Volume 1: Proposed Standards of Performance for Petroleum Refinery Sulfur Recovery Plants. EPA 450/2-76-016-a, September 1976.

3. GPA H_2S Removal Panel-5. The Ohio and Gas Journal, September 11, 1976. pp. 88-91.

4. Letter, C. M. Tyler, SOHIO, to Don R. Goodwin, U.S. EPA, dated July 15, 1982.

5. Letter, J. C. Brocoff, Ralph M. Parsons Co., to S. T. Cuffe, U.S. EPA, February 16, 1983.

6. Memorandum - "Tail Gas Sulfur Recovery Costs", C. Sedman, U.S. EPA, March 2, 1983.

7. SO_2 Emissions in Natural Gas Production Industry - Background Information for Proposed Standards - Draft EIS. EPA-450/3-82-023a, January 1983, pp. 8-5 and 8-6.

8. Smith, J. M. and Van Ness, A.C. Introduction to Chemical Engineering Thermodynamics, Section 5.8, pp. 147-153, McGraw-Hill, 1959.

9. Control of Emissions from Lurgi Gasification Plants, Appendix C, pg. C-19, EPA-450/2-78-012, March 1978.

10. Reference 5.

11. Letter, H. J. Grimes, ARCO Petroleum Products Co. to C. Sedman, U.S. EPA, dated October 5, 1982.

12. Reference 1.

13. Telephone Conversation, C. B. Sedman, U.S. EPA, and R. E. Warner, Ralph M. Parsons Co., February 1, 1983.

14. Telephone Conversation, C. B. Sedman, U.S. EPA, and T. Wohlforth, SOHIO, September 8, 1982.

15. Reference 13.

16. Confidential letter, B. F. Ballard, Phillips Petroleum, to Don Goodwin, U.S. EPA, dated July 13, 1982.

17. Telephone Conversation - C. B. Sedman, U.S. EPA and L. Landrum, ARCO, October 25, 1982.

18. Reference 5.

19. Reference 16.

20. Reference 5.

APPENDIX B
RESULTS OF COST ANALYSES FOR INTERMEDIATE CONTROL SYSTEM

As a basis for comparison of an NSPS control system analyzed in Appendix A, a lower capital cost system with control efficiency somewhere between that of a Claus and a Claus + reduction tail gas system is evaluated in this Appendix. Currently, the only available system operating in the United States and, hence a source of operating data, is the IFP-1500 system. At present, it operates at four refineries of 100, 180, 250, and 400 LT/day capacities each.[1] From these sources, operating data were obtained to enable a rough cost estimate for a 100 LT/D case as follows:

B.1 CAPITAL COST ESTIMATES

The 100 LT/D Claus plant from Figure A-2 is \$4.50 x 10^6. The incinerator from Figure A-4 is \$0.41 x 10^6. From Figure A-9, the stack cost is estimated at \$0.45 x 10^6 based on a 250 lb/hr SO_2 emission rate (98.66 percent sulfur recovery - see Reference 1). The heat recovery system is identical to that of Case 3A at \$0.56 x 10^6.

The IFP-1500 at 100 LT/D is reported to cost \$1.234 x 10^6 for a 100 LT/D system and \$2.35 x 10^6 for 180 LT/D, December 1975 basis.[2] However, the 180 LT/D was a retrofit application. Therefore, the \$1.234 x 10^6 corrected to July 1982 is approximately \$2.12 x 10^6 for the IFP portion of the Claus plant.

To make the system truly comparable to the cases examined in Appendix A, a heat recovery boiler is also required, estimated at \$0.56 x 10^6. Therefore, the total investment is \$8.04 x 10^6 for a 3-stage 100 LT/D Claus plant with IFP-1500 tail gas treatment, incinerator with waste heat recovery, and stack.

B.2 OPERATING COST ESTIMATES

All Claus operating costs will be taken by procedures in Appendix A, in most instances transformed directly from Case 3A. Fuel gas requirements for the incinerator, however, must be recalculated due to inlet gas temperature differences. For simplicity, it is assumed that the steam generation by Claus stages are identical to Case 3A, although in actual practice, the first stage might be operated at higher temperatures (less net 250 # steam generation, more 50 # steam generation) than in Claus only operation, in order to minimize sulfide formation with carbon dioxide (COS + CS_2).

For IFP operating costs, the following estimates for a 100 LT/D unit are used based upon letters from operating facilities:[3,4]

- utility requirements:
    - electricity 21 KWH/H
    - condensate 1.5 gpm
- chemical/catalyst requirements (include routine make-up and periodic inventory replacement)
    - solvent (PEG + salicyclic acid + sodium hydroxide): 124,000 lb/yr

These figures are based on an assumed solvent inventory of 62 short tons with 50 percent replacement annually and a complete inventory replacement every two years; equivalent to a 62 short ton replacement annually. Again, this is a simplification as the sodium hydroxide and salicyclic acid are replaced more frequently than the polyethylene glycol (PEG), but are minor (1 percent each) components of the overall solvent. PEG costs in 1982 varied from $.46/lb Gulf Coast to $.53/lb West Coast, so an average of $0.50/lb is used.[5,6]

Maintenance costs are assumed as an annual 3.55 percent of the IFP capital cost. Two plants surveyed reported costs at 3.41 and 3.74 percent, respectively.[7,8]

All other costs are assumed similar to those in Appendix A and are calculated as a function of capital and operating costs accordingly.

B.3 LINE ITEM COSTS

Case 3C 100 LT/D (80% H_2S)

Item	Source	Estimate	
		Claus	IFP
Capital Cost:			
3-Stage Claus	Figure A-2	$4.50 $\times 10^6$	
IFP	Section B.1		$2.12 $\times 10^6$
Incinerator	Figure A-4	$0.41 $\times 10^6$	
Stack	Figure A-9	$0.45 $\times 10^6$	
Waste Heat Recovery	Figure A-5	$0.56 $\times 10^6$	
		$5.92 $\times 10^6$	
Operating Cost (credit)			
600 psig	Table A-8	(4600 lb/hr)	--
250 psig	"	(15740 lb/hr)	--
50 psig	"	(6040 lb/hr)	--
15 psig	"	(1240 lb/hr)	--
condensate	Table A-8/Section B.2	(3420 lb/hr)	750 lb/hr
electric power	Figure A-14/Section B.2	212 KWH/H	21 KWH/H
fuel gas	calculated	6.15 $\times 10^6$ Btu/hr	
catalyst	Section A.2.2	2,084 lb/yr	
solvent	Section B.2	124,000 lb/yr	

The corresponding costs are tabulated in Table B-1 and compared to the Claus only case (3A) in Table B-2.

Table B-1. LINE ITEM COST FOR CASE 3C

Capital Cost - $\$8.04 \times 10^6$

Direct Operating Cost	i = 15%	i = 10%	i = 20%
A. Utilities and Chemicals			
1. treated boiler feedwater	392,850	392,850	392,850
2. electric power	97,860	97,860	97,860
3. fuel gas	180,810	180,810	180,810
4. catalyst	8,010	8,010	8,010
5. solvent	62,000	62,000	62,000
B. Labor			
1. operators	169,360	169,360	169,360
2. supervision	82,340	82,340	82,340
C. Maintenance & Repair	252,860	252,860	252,860
D. Supplies and Lab Charges	33,870	33,870	33,870
Fixed Charges			
A. Capital	1,375,320	1,057,100	1,719,110
B. Taxes	80,400	80,400	80,400
C. Insurance	48,240	48,240	48,240
Plant Overhead	105,550	105,550	105,550
General Expenses			
A. Administrative	28,895	25,715	32,340
B. Distribution and Sales	28,895	27,715	32,340
Total Annualized Costs	2,947,260	2,622,680	3,297,940
Credits			
1. 600 psig steam	280,140	280,140	280,140
2. 250 psig steam	892,460	892,460	892,460
3. 50 psig steam	291,730	291,730	291,730
4. 15 psig steam	46,870	46,870	46,870
5. steam condensate	28,035	28,035	28,035
6. sulfur	4,143,720	4,143,720	4,143,720
Total Credits	5,682,955	5,682,955	5,682,955
Net Annual Operating Cost for Case 3C	($2,735,695)	($3,060,275)	($2,385,015)

Table B-2. COST-EFFECTIVENESS OF IFP CONTROL

i = 10 percent

Base Case Annual Cost, $	($3,566,420)
Base Case SO_2 Removed, tons/yr	75734.4
Claus + IFP Annual Cost, $	($3,060,275)
Claus + IFP SO_2 Removed, tons/yr	77349.44
Cost Effectiveness, $/ton	$313

i = 15 percent

Base Case Annual Cost, $	($3,313,660)
Base Case SO_2 Removed, tons/yr	75734.4
Claus + IFP Annual Cost, $	($2,735,695)
Claus + IFP SO_2 Removed, tons/yr	77349.44
Cost-Effectiveness, $/ton	$358

i = 20 percent

Base Case Annual Cost, $	($3,040,630)
Base Case SO_2 Removed, tons/yr	75734.4
Claus + IFP Annual Cost, $	($2,385,015)
Claus + IFP SO_2 Removed, tons/yr	77349.44
Cost-Effectiveness, $/ton	$406

B.4. REFERENCES

1. "Survey Report on SO_2 Control Systems for Non-Utility Combustion and Process Sources - May 1977", prepared by PEDCo Environmental, Inc., Contract No. 68-02-2603.

2. Telephone Conversation, C. B. Sedman, U.S. EPA, and B. F. Ballard, Phillips Petroleum Co., dated December 2, 1982.

3. Confidential Letter, C. Rice, Amoco, to C. Sedman, U.S. EPA, dated October 18, 1982.

4. Confidential Letter, J. E. Hardaway, TOSCO, to C. Sedman, U.S. EPA, dated January 14, 1983.

5. Reference 3.

6. Reference 4.

7. Reference 3.

8. Reference 4.